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PRELIMINARY DESIGN AND SIMULATION OF CONVENTIONAL WASTEWATER RENOVATION SYSTEMS USING THE DIGITAL COMPUTER

Robert Smith

Federal Water Pollution Control Administration
Cincinnati, Ohio

March 1968

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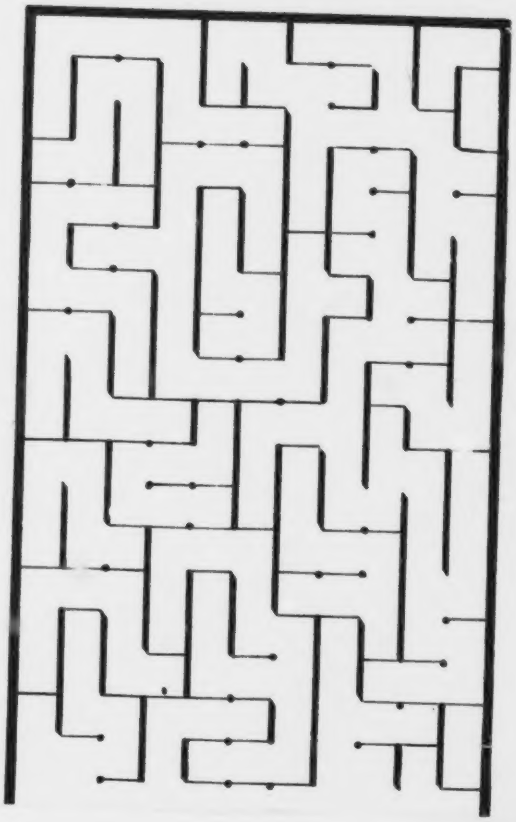

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PRELIMINARY DESIGN AND SIMULATION OF CONVENTIONAL
WASTEWATER RENOVATION SYSTEMS
USING
THE DIGITAL COMPUTER

Robert Smith

U. S. DEPARTMENT OF THE INTERIOR
Federal Water Pollution Control Administration
Advanced Waste Treatment Branch, Division of Research
Cincinnati Water Research Laboratory
Cincinnati, Ohio
March, 1968

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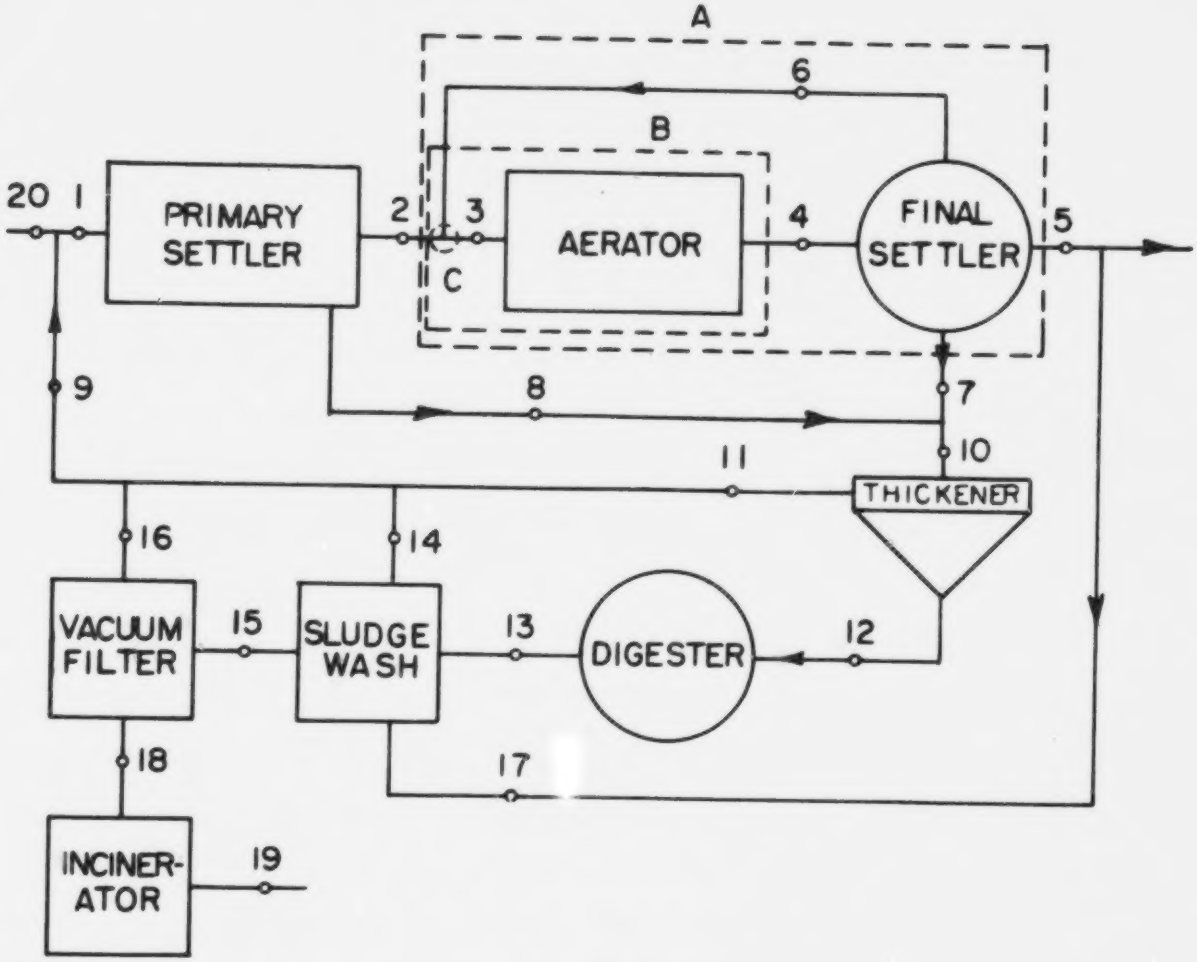
INTRODUCTION

Over the past 50 years extensive research has been conducted by universities, equipment manufacturers, and governmental agencies to delineate the physical, chemical, and biological relationships which govern the operation of wastewater treatment processes. The literature on the performance of the individual processes is voluminous but the techniques for calculating the performance of groups of processes working together as a system are limited. This is particularly true when the performance of the system as a whole is to be computed over a wide range of design and operating decisions associated with the individual processes. This report is an effort to bring together in one computational scheme the significant cost and performance relationships for the group of processes shown in Figure 1 and to attempt to calculate the performance and cost of the system as a whole, based on relationships which have been developed for the processes individually.

This requires making some bold assumptions about points which are usually not given adequate attention by researchers. These include the rate of synthesis of dissolved biodegradable carbon as compared to the rate of synthesis of particulate biodegradable carbon, or the settleability of each of the several classes of solids held in the aerator as compared to the settleability of the solids as a whole. Assumptions made will be pointed out in the discussions of mathematical models adopted for each individual process in the following pages.

In this report the term stream vector is used to designate a list of numerical values which completely describe or specify the stream. For example, the stream vector would include the stream identification


A
6
B
20
1
PRIMARY SETTLER
2
3
C
AERATOR
4
FINAL SETTLER
5
9
8
7
10
11
THICKENER
16
14
VACUUM FILTER
15
SLUDGE WASH
13
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12
18
17
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CONVENTIONAL PROCESSES SYSTEM DIAGRAM

Figure 1

number, the volume flow (mgd), the temperature, and the concentrations of each of the 11 contaminants considered in this mathematical model. Similarly, the decision vector is a list of numerical values which specify the mode of operation or the design criteria which apply to each individual process. For example, the decision vector for the primary settler contains two parameters: the fraction of suspended solids removed in the process (FRPS), and the concentration of suspended solids in the underflow stream expressed as URPS.

To calculate the performance of a process means to calculate the stream vectors for all effluent streams when the stream vectors for all influent streams and the decision vector for the process are given. As part of this calculation, the size of the structures and equipment associated with the process are computed. The amount of chemicals, electrical power, digester gas, or other minor streams directly associated with operation of the process are also computed.

A diagram of the system of processes considered in this report is shown in Figure 1. The stream vector at station 20 is supplied as input to the computational scheme. Effluent streams from upstream processes are then used as influent streams to downstream processes and the performance of each process in the system is computed in turn. When feedback streams appear, an iterative routine is required to solve for the streams in the feedback loop. For example, stream 9 would be a feedback stream if it were directed back to the head of the plant.

When the performance of the entire system has been found, the capital cost, the operating cost, and the cost of chemicals, power, and other consumable items are computed for the processes individually and summed to find the total system cost. We, of course, expect the total cost to increase monotonically with the level of performance achieved.

A determined effort has been made to use the most valid physical, chemical, and biological relationships available and to consult with prominent engineers in the sanitary engineering field, such as Dr. Perry L. McCarty[3] of Stanford University and Mr. W. W. Eckenfelder[1] of the University of Texas. In spite of this, it is recognized that many assumptions have been made which are not adequately supported by engineering and scientific investigation. Studies are being initiated by the Federal Water Pollution Control Administration (FWPCA) to supply the more critically needed data. Emphasis has been placed on the use of full-sized plant data wherever possible.

For some processes, such as the thickener, adequate relationships are not available to relate the required size of thickener to the degree of thickening specified. In cases such as this, standard loading parameters are used and no compromise is possible between cost and performance.

The computational scheme described in this report has been programmed for the digital computer using the FORTRAN IV language, which is the most recent version of FORTRAN. The program listing, together with a sample printout, is shown in the Appendix. Definitions of symbols used in the text and in the program are shown in Table A-1 of the Appendix.

The digital program contains logic to return stream 9 to the head end of the plant or to send it outside of the system. The system has been computed using both alternatives but the results of the computations shown in this report are all based on returning stream 9 to the plant.

The program will be used first to interpret and correlate new data and ultimately as a tool to raise the scientific level of the preliminary design phase of planning for wastewater treatment systems. It will be possible to study the cost-performance compromises available and finally to optimize the conventional system for any degree of treatment required.

CLASSIFICATION OF CONTAMINANTS

The scheme selected for classification of contaminants in wastewater was dictated first by the fact that the system of processes shown in Figure 1 is used primarily to treat domestic sewage. The second consideration was to find a classification which reflects the criteria now used, or which might be adopted in the future, for specifying water quality. For example, BOD, nitrogen, and phosphorus concentrations are used as measures of water quality for discharge into streams or lakes. The concentration of carbon has been suggested as a criterion for potable water quality. Many contaminants used as indicators of water quality such as heavy metals, pesticides, or coliforms have not been included. It is hoped that these might be included in the program at some time in the future.

Finally, the selection of the classification scheme was influenced by the requirement that any computed effluent stream vector must be adequate as an influent stream vector for any downstream process which might be selected for use in the system. For example, the performance of most processes is determined, in part, by the nature of the contaminant species present in the influent stream. Since many processes are concerned with separation of dissolved and particulate phases, it is necessary to retain the distinction between solid and dissolved phases. Because only part of the carbon present is susceptible to biodegradation in the environment provided by the processes, carbon is classified as biodegradable or non-biodegradable.

The classification scheme adopted, which is shown in the following table, is believed to be a good compromise between what can be handled by the present "state of the art" and what might be necessary ultimately as the performance characteristics of the processes become better understood.

	Carbon				
	Non-degradable	Degradable	Nitrogen	Phosphorus	Inorganic Fixed Matter
Solid	SNBC	SOC - SNBC	SON	SOP	SFM
Dissolved	DNBC	DOC - DNBC	DN	DP	DFM

The primary classification of the contaminants present in the various streams was taken as chemical nature, that is, carbon, nitrogen, and phosphorus. The carbon was further divided into biodegradable and non-biodegradable fractions giving four major classifications. A fifth major classification was added for inert inorganic material (fixed matter). Each of the five major classes mentioned was further divided into solid and dissolved fractions, giving a total of 10 classes. Alkalinity, not shown above, was computed for streams leading into and out of the digester.

The dissolved nitrogen is primarily ammonia nitrogen, and the dissolved phosphorus primarily orthophosphate (PO_4). The units for all contaminant species are milligrams per liter of carbon, nitrogen, phosphorus, and fixed matter. The units are changed to lb/day for stations 18 and 19.

In order to make maximum use of process performance data which are expressed in terms of standardized tests*, such as 5-day BOD, TOC. COD, VSS, and TSS, it is necessary to use conversion factors. The following table shows the conversion factors used and the reference where the measurement was reported.

Conversion Ratio	Raw Sewage	Activated Sludge	Reference
BOD/TOC	1.87		1, 4
COD/VSS	1.50	1.42	5, 7, 24
VSS/TOC	(2.13)	(1.90)	
BOD/VSS	(0.80)	0.84	24
COD/TOC	3.2	2.7	4, 7

Conversion ratios shown in parentheses have been derived from one or more of the measured ratios.

---

*BOD = Biochemical Oxygen Demand
TOC = Total Organic Carbon, mg/1 of carbon
COD = Chemical Oxygen Demand
VSS = Volatile Suspended Solids
TSS = Total Suspended Solids

COMPOSITION OF DOMESTIC SEWAGE

The A.S.C.E. Manual No. 36 "Sewage Treatment Plant Design"[13] contains data on the removal of BOD and suspended solids by the primary settler for 57 plants in the United States. These data were used to find an average BOD value for raw sewage (197.2 mg/l) and to estimate the fraction of the BOD associated with the suspended solids. For example, let $f_1 = TSS_2/TSS_1$. We then assume that $SBOD_2/SBOD_1$ equals f_1. The concentration of dissolved BOD at station 2 ($DBOD_2$) is set equal to the dissolved BOD at station 1. The ratio of total BOD at station 2 to the total BOD at station 1 is called f_2 and expressed as follows:

$$f_2 = \frac{SBOD_2 + DBOD_2}{SBOD_1 + DBOD_1}$$

Letting a equal the ratio of solid BOD to total BOD at station 1, we can then express a as follows:

$$a - (1 - f_2)/(1 - f_1)$$

Since f_1 and f_2 are given in Manual No. 36 for each of the plants, a was computed and averaged. The average value was 0.715 indicating that the BOD of raw sewage is about 70% settleable and 30% dissolved.

For use in the computer program the biodegradable carbon at station 20 was set at 32 mg/l of dissolved carbon and 75 mg/l solid carbon. The 5-day BOD at station 20 was then computed as 59.84 mg/l of dissolved BOD and 140.3 mg/l of solid BOD. Measurements of dissolved non-biodegradable organics in the effluent from extended aeration plants and in well-operated activated sludge plants suggest that the dissolved non-biodegradable carbon averages about 11 mg/l carbon. This value was used in the program.

McKinney[25] states that solid non-biodegradable organic matter in raw sewage is about 50 mg/l mass, which corresponds to about 21.5 mg/l carbon.

The removal of carbon in the digester as reported by McCarty[52] shows that about 60% of the carbon is removed with a digester detention time of 15 days. In order to make the program duplicate this observation, it was necessary to set the solid non-biodegradable carbon at station 20 at 30 mg/l carbon.

Painter[5] measured the ratio of organic nitrogen to organic carbon in raw sewage and found a value of about 0.10 for both dissolved and particulate fractions. The organic nitrogen concentrations at station 20 were, therefore, taken as 10% of the total carbon for both solid and dissolved fractions. In a survey of 18 plants made by the FWPCA[8] an average concentration of 14.8 mg/l of ammonia nitrogen was found. This, added to the 4.3 mg/l of organic dissolved nitrogen, gives a total of 19.1 mg/l of dissolved nitrogen.

Measurements made by Rudolfs[10] on the concentration of phosphorus in raw sewage averaged to 2.3 mg/l of phosphorus, with 67% being in the dissolved form. Levin[22] measured the total phosphorus in raw sewage at the Washington, D.C. plant and found an average of 5 mg/l. A 31-plant survey made by FWPCA[8] measured inorganic dissolved phosphorus and found an average of 8.2 mg/l. Sawyer[12] states that the inorganic phosphorus content of sewage is in the range of 2 to 3 mg/l, while the organic forms total from 0.5 to 1.0 mg/l. For use in the program it was assumed that dissolved phosphorus was 4 mg/l and solid phosphorus was 2 mg/l.

Solid fixed matter was measured in raw sewage by Painter[9] and the average value was 136 mg/l in sewage with a BOD of about 600 mg/l.

Heukelekian[6] found an average of 25 mg/l of fixed solid material in sewage at four different New Jersey plants. In the program, solid fixed matter at station 20 was set at 30 mg/l mass. Dissolved fixed matter was found to be in the range 500 to 700 mg/l from a survey made by FWPCA[8]. To summarize, the following table shows the assumed concentrations for all contaminants at station 20 used in the program:

	Carbon				
	Non-degradable	Degradable	Nitrogen	Phosphorus	Inorganic Fixed Matter
Solid	30.0	75.0	10.0	2.0	30.0
Dissolved	11.0	32.0	19.1	4.0	500.0

PRIMARY SETTLER

The performance of the primary settler is specified by two parameters. The first, FRPS, is the fraction of the solids entering the settler at station 1, which are removed from the main stream and sent to the thickener at station 8. The second parameter, URPS, expresses the degree of thickening accomplished in the primary settler. These parameters are defined as follows:

$$\text{FRPS} = (Q_8 \text{SOC}_8)/(Q_1 \text{SOC}_1) \tag{1}$$

$$\text{URPS} = \text{SOC}_8/\text{SOC}_1 \tag{2}$$

FRPS and URPS are supplied as input to the program. The fractional removal of suspended solids, FRPS, can be related to the size of settler required, but no relationship is available for relating URPS to the design parameters of the settler, such as the sludge storage volume provided or the sludge pumping rate.

By writing material balances around the settler, the following equations can be derived for the flow and concentration of carbon in the form of suspended solids at stations 2 and 8.

$$Q_2 = Q_1 - Q_8 \qquad (3)$$

$$Q_8 = Q_1(FRPS/URPS) \qquad (4)$$

$$SOC_8 = SOC_1\ FRPS\ Q_1/Q_8 \qquad (5)$$

$$SOC_2 = SOC_1(1 - FRPS)Q_1/Q_2 \qquad (6)$$

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It is assumed that the removal of all other species of suspended solids can be calculated in a similar way. That is, a single value of FRPS or URPS applies to all classes of suspended solids. This assumption has never been tested by measurements on the streams into and out of a primary settler, but it has been suggested for use by some investigators.

The primary settler overflow rate required to accomplish the demand removal of suspended solids (FRPS) is expressed as follows:

$$FRPS = 0.82\ e^{-GPS/2780} \qquad (7)$$

This relationship was derived by plotting data for suspended solids removal versus overflow rate for full-sized plants, as reported in "Sewage Treatment Plant Design."[13]

It is assumed that no dissolved species are removed in the primary settler.

AERATOR AND FINAL SETTLER

The 5-day BOD to be achieved at station 5 (BOD_5) and the mixed liquor suspended solids (MLSS) to be carried in the aerator are supplied to the program as input. The size of the aerator required is calculated, using the following relationship supplied by Eckenfelder[1]:

$$VAER = \frac{Q_2(BOD_2 - BOD_5)}{CAER \cdot MLSS \cdot BOD_5} \qquad (8)$$

This relationship can be derived by assuming that the BOD which leaves the aerator at stations 5 and 7 is dissolved. The following material balance can then be written for control volume "A", shown in Figure 1:

$$Q_2 \cdot BOD_2 = Q_2 \cdot BOD_5 + CAER \cdot VAER \cdot MLSS \cdot BOD_5 \qquad (9)$$

This can be rearranged to give equation 8. The rate constant CAER is temperature-dependent, and the following relationship is recommended by Eckenfelder:

$$CAER = CAER_{20}(1.047)^{T-20} \qquad (10)$$

Using equation 8 to fit data from full-sized activated sludge plants, Eckenfelder found the value of $CAER_{20}$ to be 0.024. Equation 8 is used in the program to find the volume of the aerator, in keeping with the philosophy applied throughout the program of using full-sized plant data wherever possible.

After sizing the aerator, using a relationship derived from full-sized plant data, the next problem is to make an effort to compute the flow and the concentration of all dissolved and solid species at stations 5 and 7, based on known governing principles for the activated sludge process.

McCarty[24] studied measurements made by several investigators and concluded that for every pound of 5-day BOD which enters the aerator and is synthesized to active solids (MLASS), 0.65 pound of active solids is produced. Some of the active solids produced are destroyed by natural causes in the aerator, and 81.5% of the mass of the active solids destroyed is converted to carbon dioxide and water as a result of endogenous respiration. Eighteen and one-half percent of the weight of the destroyed active solids is non-biodegradable and, therefore, tends to accumulate in the aerator as MLDSS. The fraction of the active solids destroyed each day (CEDR) was found by McCarty[3] to depend on the water temperature in the following way:

$$\text{CEDR} = 0.18(1.047)^{T-28} \qquad (11)$$

A key problem in computing the amount of various classes of solids at stations 5 and 7 is the differing rate of synthesis for the solid and dissolved fractions of the BOD entering at station 2. These rates cannot be approximated, although it is known that soluble food is more readily available for synthesis than solid food. It has, therefore, been assumed that in achieving the demand BOD at station 5, all of the soluble food is used first. The solid food is used only after all of the soluble food has been synthesized, and it is assumed that the remaining solid food is broken down into 70% solid and 30% dissolved.

An iterative scheme, to be discussed more fully later, is used to find the concentration of active solids (MLASS) in the aerator, and the amount of synthesis which occurs in the aerator each day. First, it must be recognized that the BOD measured at station 5 is a composite of (1) the active solids which escape over the final weirs, and (2) a part of the biodegradable organics (both solid and dissolved) entering at station 2 which have not been used as food in the aerator.

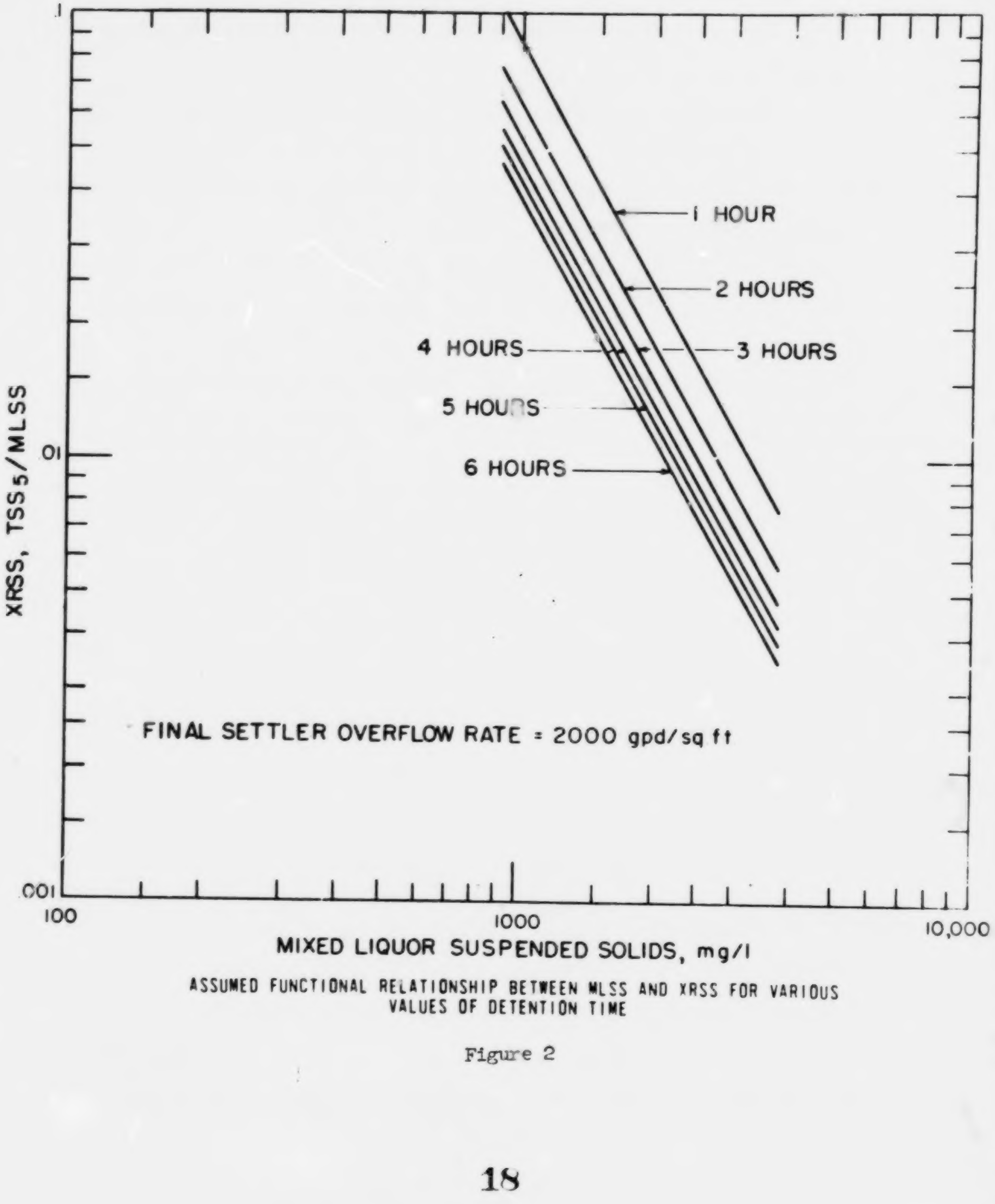
Before any mass balances can be written around control volume "A" (Figure 1), we must specify the performance of the final settler by two parameters. The first, XRSS, is the fraction of the solids held in the aerator (MLSS) which escapes over the final settler weirs. Bench scale experiments made by Villiers[35] at the Cincinnati Water Research Laboratory, using fill and draw activated sludge methods and settling columns, have produced the following empirical relationship between XRSS and the design settler overflow rate, GSS, the aerator mixed liquor suspended solids, MLSS, and the detention time, TA, days, for the aerator:

$$\mathrm{XRSS} = \frac{556(\mathrm{GSS})^{0.494}}{\mathrm{MLSS}^{1.82}(24 \cdot \mathrm{TA})^{0.439}} \qquad (12)$$

A plot of this relationship is shown in Figure 2. The second parameter, URSS, is defined as the ratio of the solids concentration in the final settler underflow (station 7) to the solids concentration (MLSS) in the aerator. Since URSS can be controlled by the underflow pumping rate, this parameter is given a value of 3.0 in the program. The upper limit for URSS is set at 1,200,000/MLSS/SVI. SVI is the sludge volume index which is taken as 100.

GSS, the overflow rate for the final settler in gpd/sq ft, is supplied to the program as input. It should be noted here that the same values for XRSS and URSS are applied to every class of solid material held in the aerator. This is probably a significant weakness in the theory as it now exists.

Even though, at this point, we do not have values for MLASS or the amount of BOD synthesized into active solids each day (expressed below



ASSUMED FUNCTIONAL RELATIONSHIP BETWEEN MLSS AND XRSS FOR VARIOUS VALUES OF DETENTION TIME

Figure 2

as $Q_2 \cdot FOOD$), we can write the following material balance for active solids around control volume "A".

$$0.65FOOD \cdot Q_2 - CEDR \cdot VAER \cdot MLASS = MLASS(Q_5 \cdot XRSS + Q_7 \cdot URSS) \qquad (13)$$

This material balance equation can be solved for the volume of the activated sludge waste stream as follows:

$$Q_7 = \frac{Q_2 \left[\frac{0.65FOOD}{MLASS} - XRSS \right] - CEDR \cdot VAER}{URSS - XRSS} \qquad (14)$$

The concentration of non-biodegradable material originating from the destruction of active sludge in the aerator (MLDSS) can now be found by writing the following material balance around control volume "A".

$$0.12FOOD \cdot Q_2 = (MLDSS + 0.185MLASS) \cdot (XRSS \cdot Q_5 + URSS \cdot Q_7) \qquad (15)$$

The following relationship is thus derived from MLDSS:

$$MLDSS = \left[\frac{0.12FOOD \cdot Q_2}{Q_5 \cdot XRSS + Q_7 \cdot URSS} \right] - 0.185MLASS \qquad (16)$$

The part of the influent (station 2) biodegradable solids which is not synthesized to active solids, and thus forms part of the mixed liquor suspended solids, is designated as MLBSS and can be calculated from a material balance around control volume "A" as follows:

$$MLBSS = \frac{Q_2 \cdot SBOD_4}{(Q_5 \cdot XRSS + Q_7 \cdot URSS)0.80} \qquad (17)$$

$SBOD_4$ is the part of $SBOD_2$ which has not been synthesized.

The concentration of inert inorganic solids in the aerator (MLISS), and the concentration of non-biodegradable organic solids (MLNBSS) resulting from the non-biodegradable organic solids entering the aerator at station 2, can be found by writing similar mass balances as follows:

$$\mathrm{MLISS} = \frac{Q_2 \cdot \mathrm{SFM}_2}{Q_5 \cdot \mathrm{XRSS} + Q_7 \cdot \mathrm{URSS}} \qquad (18)$$

$$\mathrm{MLNBSS} = \frac{Q_2 \cdot \mathrm{SNBC}_2}{Q_5 \cdot \mathrm{XRSS} + Q_7 \cdot \mathrm{URSS}} \qquad (19)$$

The iterative scheme used to find the FOOD synthesized each day and the active solids concentration in the aerator (MLASS) is based on Bolzano's method[31] for root finding, and is arranged so that the iterative loop to find FOOD is within the iterative loop to find the correct value for MLASS. To begin, we know that MLASS has a value between zero and some maximum value. The maximum value for MLASS is taken as the greater of two values: The input value of MLSS, or the value of MLASS which would make BOD_5 equal the demand BOD_5 if all the entering food is synthesized. The first guess is, therefore, at the mid-point of this range. Having selected a value for MLASS, we next compute the limits of the range for FOOD as follows:

$$\mathrm{FOOD}_{max} = \mathrm{SBOD}_2 + \mathrm{DBOD}_2 \qquad (20)$$

$$\mathrm{FOOD}_{min} = (\mathrm{CEDR} \cdot \mathrm{VAER}/Q_2 + \mathrm{XRSS}) \cdot \mathrm{MLASS}/0.65 \qquad (21)$$

FOOD_{max} is the total food available for synthesis entering at station 2. FOOD_{min} is found by setting the waste stream, Q_7, equal to zero because, having selected a value for MLASS, the amount of active solids synthesized must at least equal the amount destroyed by natural causes, plus the amount which escapes in the effluent at station 5. The starting value for FOOD is taken at the mid-point of the range.

Having selected values for MLASS and FOOD we can now use equations 14, 16. 17, 18 and 19 to compute the concentration of all contaminant species at stations 5 and 7. FOOD is adjusted, using the Bolzano bisection method. until the concentration of BOD at station 5 equals the demand value within some tolerance. The tolerance used in the program is 0.10 mg/1. When the iterative loop for FOOD has been satisfied, the mixed liquor suspended solids concentration (MLSS) is computed as follows:

$$MLSS = MLASS + MLBSS + MLNBSS + MLDSS + MLISS$$

If this computed value for MLSS differs from the input value, a new value for MLASS is assigned by employing the Bolzano bisection method, and the iteration to make BOD_5 equal the input value is repeated with the new value for MLASS. This procedure is continued until MLSS equals the input value within a specified tolerance. A tolerance of 5 mg/1 was used for MLSS. The final value for MLASS is then taken as the concentration of active solids in the aerator.

Finally, the volume of the return activated sludge stream (Q_6) is found by writing the following material balance for active solids around control volume "B".

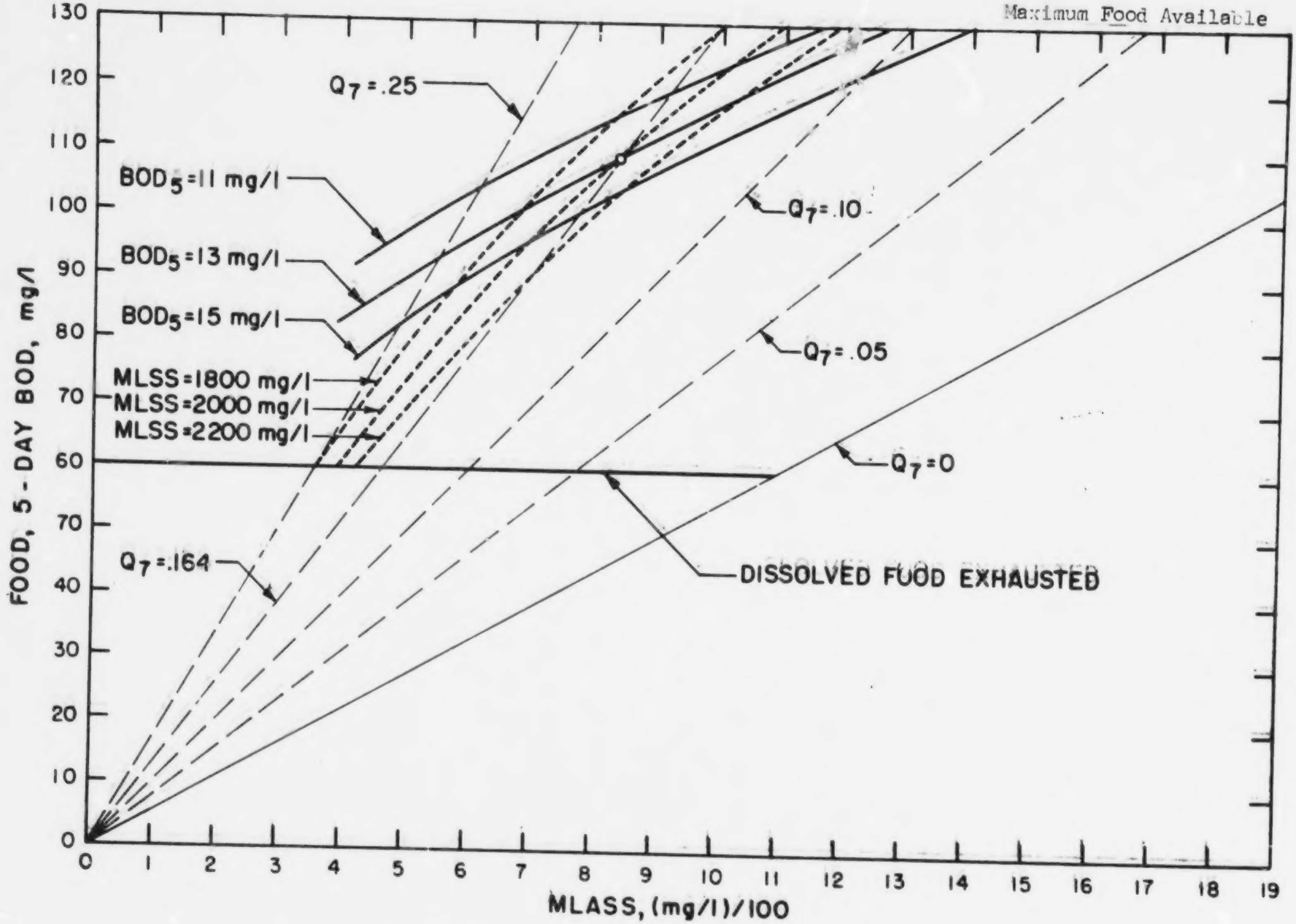
$$0.65FOOD \cdot Q_2 + Q_6 \cdot URSS \cdot MLASS = Q_4 \cdot MLASS + CEDR \cdot VAER \cdot MLASS \qquad (22)$$

Since Q_4 is the sum of Q_2 and Q_6, equation 15 can be solved for Q_6 as follows:

$$Q_6 = \frac{Q_2\left[1.0 - \frac{0.65FOOD}{MLASS}\right] + CEDR \cdot VAER}{URSS - 1.0} \qquad (23)$$

Porges, et al.,[26] measured the nitrogen content of active cells and found it to be (0.234) times the carbon content. The nitrogen content of all other particulate organic compounds at stations 5 and was taken as (0.10) times the carbon content. From these two assumptions the solid organic nitrogen was computed at stations 5 and 7. The remainder of the nitrogen was assumed to be dissolved. Solid organic phosphorus was taken as 1% of the particulate carbon at stations 5 and 7. The remaining phosphorus was assumed to be dissolved.

The iterative scheme used in the program is illustrated by Figure 3. This chart was drawn for the case where MLSS was set equal to 2000 mg/l and BOD_5 was set equal to 13 mg/l. The purpose of the iterative scheme is to find the point in the FOOD-MLASS domain where MLSS equals 2000 and BOD_5 equals 13. The lines of constant Q_7 and $SBOD_4$ can be easily drawn from equation 14 and the assumption that remaining solid FOOD is broken down into 70% solid and 30% dissolved. Notice that the feasible domain is bounded by the coordinate axes, the $Q_7 = 0$ line and the line representing full utilization of the available FOOD. Calculated lines of constant MLSS and constant BOD_5 are shown. Notice that the MLSS = 2000 line intersects the $BOD_5 = 13$ line when FOOD 109 mg/l and MLASS = 837 mg/l. This point was also found by the iterative technique built into the digital computer program. This hand calculation was used to check the operation of the computer program.



CONTOURS OF MLSS AND BOD_5 FOR XRSS = .0124 AND VAER = 1.872 mg

Figure 3

NITRIFICATION IN THE AERATOR

The basic factors which control the oxidation of ammonia in the activated sludge process have been investigated with great thoroughness by Downing, et al.,[38] in England.

The key concept involved in adjusting the size and operating conditions of the activated sludge plant to assure nitrification is that, for steady-state operation, the BOD-consuming bacteria must be wasted at the rate they are generated, minus the rate they are destroyed by endogenous respiration. This, necessarily, controls the wasting rate of nitrifying bacteria. If the wasting rate for nitrifying bacteria exceeds the rate at which they are generated, no nitrification can occur. If the nitrifying population can be maintained, however, Downing shows how the extent of nitrification can be calculated. It has been shown that the nitrogen required for the synthesis of new sludge is approximately equal to the organic nitrogen available from the food. The ammonia to be oxidized is, therefore, assumed equal to the influent ammonia. Downing also points out that a particular organism (Nitrosomonas) is responsible for conversion of ammonia to nitrite, and a second organism (Nitrobacter) converts nitrite to nitrate. Experience has shown, however, that when conditions are favorable for oxidizing ammonia, the ammonia oxidized is converted almost wholly to nitrate. To promote nitrification the emphasis is, therefore, placed on maintaining a population of Nitrosomonas. Since all of the design and operating conditions for the aerator have been set by the demands for BOD removal, the question we must answer is whether conversion of the ammonia to nitrate will occur. It is always assumed that sufficient oxygen is supplied, that no inhibitory substances are present, and that the dissolved oxygen concentration is sufficiently high. Downing states that 0.5 ppm ± 0.2 ppm of dissolved oxygen is sufficient for sustaining nitrification. If we let the concentration of *Nitrosomonas* be designated by X, we can write the following material balance for *Nitrosomonas* around control volume "C", where p is the ratio of Q_6 to Q_2:

$$p\cdot Q_2\cdot X_4\cdot \mathrm{URSS} = (1 + p)Q_2\cdot X_3 \tag{24}$$

The ratio of *Nitrosomonas* concentration at station 3 and 4 is, therefore, given as follows:

$$X_4/X_3 = (1 + p)/(p\cdot \mathrm{URSS}) \tag{25}$$

Each element of water is assumed to move down the aeration tank without mixing. The concentration of *Nitrosomonas* (X) and the concentration of ammonia (S) are calculated as a function of time of travel (t) by integrating the following two differential equations:

$$\frac{dX}{dt} = \frac{\mu XS}{K + S} \tag{26}$$

$$\frac{dX}{dt} = Y\,\frac{dS}{dt} \tag{27}$$

t is measured in days. Equation (27) can be integrated to give the following expression:

$$X = X_3 + Y(S_3 - S) \tag{28}$$

The values given by Downing[38] for the constants are $\mu = 0.33$ (days)$^{-1}$, K = 1.0 mg/l *Nitrosomonas*, and Y = .05 gm cells/gm of NH_3-N used. By substituting (28) into (26), an expression is obtained which can be integrated to yield the following equation:

$$\left[\frac{Y(K + S_3) + X_3}{X_3 + YS_3}\right] \ln(X_4/X_3) - \left[\frac{KY}{X_3 + YS_3}\right] \ln(S_4/S_3) = \mu t \qquad (29)$$

Letting X equal X_3 and S equal S_3. equation (28) becomes:

$$X_4 = X_3 + Y(S_3 - S_4) \qquad (30)$$

Equations (29) and (30) can now be used to calculate the extent of oxidation of ammonia in the plug flow aerator. Dividing both numerator and demoninator of the coefficients of the first two terms of equation (29) by S_3Y, we obtain the following equation:

$$\left[\frac{(K/S_3 + 1) + X_3/(S_3Y)}{1 + X_3/(S_3Y)}\right] \ln\,(X_4/X_3) - \left[\frac{K/S_3}{X_3/(S_3Y) + 1}\right] \ln\,(S_4/S_3) = \mu t \qquad (31)$$

Since Y is given a value of 0.05 and K a value of 1.0 by Downing, the coefficient of the first term of (31) is approximately equal to 1.0. Equation 31 can, therefore, be written more simply as follows:

$$\ln\,(X_4/X_3) - \left[1/(X_3/Y + S_3)\right] \ln\,(S_4/S_3) = \mu t \qquad (32)$$

Now we must make an arbitrary decision about the residual ammonia remaining in the water at station 4 when nitrification has occurred. It was assumed in the program that 99% of the ammonia has been converted to nitrate. S_4/S_3, therefore, has a value of 0.01, the natural logarithm of which is -4.605. By using the yield constant Y, we can also write the following relationship:

$$X_4 - X_3 = Y(S_3 - S_4) \qquad (33)$$

If we substitute into this equation the fact that S_4 is 1% of S_3 we obtain the following expression for X_3/Y:

$$X_3/Y = 0.99S_3/(X_4/X_3 - 1) \qquad (34)$$

The detention time (t) used in equation 32 includes the return flow. This can be converted to the detention based on Q_2 by multiplying t by (1 + p).

More recent work by Downing[27] and co-workers has shown that the rate constant μ or CNIT, as it is called in the program, is highly temperature-dependent. The following relationship was found:

$$CNIT = 0.18\ e^{0.116(DEGC - 15)} \qquad (35)$$

Also, S_3 is found as follows:

$$S_3 = S_2/(1 + p) \qquad (36)$$

We can now substitute equations 25. 34, 35 and 36 into equation 32, and solve for the detention time (TAN) required to achieve nitrification. If the detention time (TA) previously computed to meet the BOD removal requirement equals or exceeds this value, we assume that nitrification has occurred and that all of the ammonia has been converted to nitrate. It has been shown by Downing and others that solution of the equations which describe the degree of nitrification occurring in the activated sludge predict a precipitous change from no nitrification to complete nitrification. This is generally observed in plant records. The air requirements for the aerator are increased to meet the demand if it is concluded that nitrification has occurred.

AIR REQUIREMENTS

The oxygen requirements for synthesis and endogenous respiration of BOD-using cells are normally given by a relationship of the following form:

lb O_2/day = a(lb of BOD synthesized/day) + b(lb of MLSS held in aerator) (37)

The values of a and b have been evaluated empirically by various investigators. An average measured value for a is about (0.5) and an average value for b is about (0.10). This relationship has the obvious disadvantage of being based on the total mixed liquor suspended solids instead of the active solids.

McCarty[3] has suggested the following more logical way of computing oxygen requirements, based on the concentration of active solids in the aerator. Since the COD of organic matter is defined as the amount of oxygen required to convert the organic matter to carbon dioxide and water, the oxygen requirement can be found by subtracting the COD of the cells formed from the COD of the organic material consumed as FOOD by the active cells.

lb O_2/day = (COD of FOOD used, lb) - (COD of cells produced, lb) (38)

The COD/BOD ratio for raw sewage has been estimated[5] conservatively as (1.5). For active cells the ratio of COD/VSS has been estimated [24] as (1.42). For relationships already used in calculating the aerator performance, the terms on the right hand side of equation can be calculated as follows:

COD of cells, lb/day = 1.42 (0.65FOOD$\cdot Q_2$ - 0.815MLASS$\cdot$CEDR$\cdot$VAER)8.33 (39)

$$\text{COD of FOOD used, lb/day} = 1.5\ (\text{FOOD}\cdot Q_2)8.33 \qquad (40)$$

Substituting equation (39) and (40) into equation (38) gives the following relationship for calculating the amount of oxygen used by the BOD-using organisms:

$$\text{lb } O_2 \text{ used per day} = 8.33(0.58\text{FOOD } Q_2 + 1.16\text{MLASS}\cdot\text{CEDR}\cdot\text{VAER}) \quad (41)$$

When the detention time to achieve nitrification (TAN) is less than or equal to the detention time required for BOD removal (TA) it is assumed that sufficient oxygen must be supplied to convert ammonia to nitrate. Downing gives the following quantitative relationship for conversion of ammonia to nitrate by the Nitrosomonas and Nitrobacter microorganisms:

$$2NH_4^+ + 4O_2 \longrightarrow 2NO_3^- + 2H_2O + 4H^+ \qquad (42)$$

Four moles of oxygen are, therefore, required to convert one mole of nitrogen to nitrate. The yield factor for Nitrosomonas is given as 0.05 lb of cells produced for each pound of nitrogen converted. Thus, a negligible amount (0.8%) of the nitrogen used as food is converted to cell material. The ratio of oxygen required per pound of nitrogen converted to nitrate is, therefore 64/14 or 4.57.

The density of air at sea level is taken as 0.075 lb per cu ft, and air is assumed to be 23.2% oxygen. Designating oxygen transfer efficiency as AEFF we can then express the amount of air which must be pumped into the aerator each day as follows:

$$\text{AIRCFD} = (\text{lb } O_2 \text{ used per day})/0.075/0.232/\text{AEFF} \qquad (43)$$

Oxygen transfer efficiency (AEFF) depends on many factors such as the type of equipment used, air flow per foot of tank length, and depth of diffuser. In the program the two factors taken into consideration are water temperature and the dissolved oxygen level maintained in the aerator. $AEFF_{20}$ is the oxygen transfer efficiency at 20°C and zero dissolved oxygen level. $AEFF_{20}$ is supplied to the program as input. AEFF is calculated as follows:

$$AEFF = AEFF_{20}([DOSAT - DO]/DOSAT)\cdot(1.02)^{DEGC\ -\ 20} \qquad (44)$$

Since one atmosphere (14.7 psi) is equivalent to 33.93 ft of water at 20°C, the compresser would have to pump against at least 6.5 psi when the diffusers are 15 ft below the water surface. To account for the additional losses across the diffuser and piping, Fair and Geyer[43] state that the required pressure can be estimated as 25% greater than the pressure due to submergence, or 8.1 psi. Theoretical power requirements for the blowers are found from the following formula for hydraulic horsepower:

$$\text{Horsepower} = \frac{SCFM\cdot\Delta P\cdot 144}{33000\cdot EFF} \qquad (45)$$

ΔP is expressed in psi, SCFM means standard cubic feet per minute and EFF is efficiency which was assumed to be 0.80. Converting the kilowatt-hours and standard cubic feet, we find that with a ΔP of 8.1 psi, 1830 scf of air can be supplied with the expenditure of one kilowatt-hour.

THICKENER

The performance of the thickener is determined by the solids recovery ratio (TRR), which is defined below, and the total solids in the underflow stream at station 12 (TSS_{12}). These two parameters, which are essentially engineering estimates of the average performance expected, are supplied as input.

$$TRR = \frac{Q_{12}TSS_{12}}{Q_{10}TSS_{10}} \tag{46}$$

The volume of streams 11 and 12 is then computed as follows:

$$Q_{12} = TRR \cdot Q_{10} \cdot TSS_{10}/TSS_{12} \tag{47}$$

$$Q_{11} = Q_{10} - Q_{12} \tag{48}$$

Total solids in the overflow stream (station 11) is then found as follows:

$$TSS_{11} = (1 - TRR) \cdot Q_{10} \cdot TSS_{10}/Q_{11} \tag{49}$$

The proportions of the various classes of solids in streams 11 and 12 are assumed to be equal to the proportions in stream 10. For example,

$$SOC_{12} = (TSS_{12}/TSS_{10}) \cdot SOC_{10} \tag{50}$$

$$SOC_{11} = (TSS_{11}/TSS_{10}) \cdot SOC_{10} \tag{51}$$

Concentrations of dissolved species in streams 11 and 12 are set equal to the concentrations in stream 10.

The thickener is sized by input values of solids loading in pounds per day per square foot of surface area (GSTH), and overflow rate in gallons per day per square foot of surface area (GTH). The greater of the two surface area values found is used as the required thickener size.

No relationship has been found to relate the desired thickener performance to the size of the thickener and the characteristics of the influent stream (station 10). It is hoped that this kind of information can be developed from studying the performance of full-sized thickeners.

DIGESTER

The relationships used to describe the digester were taken from the work of McCarty[52], who believes that, as a first approximation, it should be assumed that only material classified as biodegradable will be susceptible to biodegradation in the digester. The concentration of biodegradable carbon at station 12 was, therefore, computed as follows:

$$\mathrm{DIGC}_{12} = \mathrm{SOC}_{12} - \mathrm{SNBC}_{12} + \mathrm{DOC}_{12} - \mathrm{DNBC}_{12} \qquad (52)$$

Letting F equal the concentration of biodegradable carbon and C equal the concentration of active solids in the completely mixed digester, we can write the following mass balance for active solids around the digester:

$$\mathrm{VDIG}\,\frac{dC}{\partial t} = \mathrm{VDIG}\,\frac{K_1 CF}{K_2 + F} - b\cdot\mathrm{VDIG}\cdot C - Q_{13}\cdot C \qquad (53)$$

The term on the left of the equal sign represents the time rate of change of the mass of active organisms in the digester. The first term on the right-hand side of the equation is the Monod relationship expressing the growth of active solids. The second right-hand term is the decay of active solids and the third right-hand term is the active solids leaving the digester at station 13. Since we are looking for a steady-state expression, the left-hand term must be zero. McCarty found that the term for decay of active solids can be neglected, which reduces this mass balance expression to the following form:

$$\frac{K_1 F}{K_2 + F} = \frac{Q_{13}}{\mathrm{VDIG}} = \frac{1}{\mathrm{TD}} \qquad (54)$$

In the program, F is called DIGC13 and the detention time of the digester is supplied as input. This equation can then be readily solved for DIGD13. Notice that the concentration of biodegradable carbon in the effluent from the digester is independent of the biodegradable carbon in the influent stream and depends only on the specified detention time of the digester.

From laboratory investigations, O'Rourke and McCarty[54] found that K_1 could be represented as follows:

$$K_1 = 0.28\ e^{-0.036(35 - T)} \tag{55}$$

In the same study K_2 was found to depend on temperature in the following way:

$$K_2 = 700\ e^{+0.10(35 - T)} \tag{56}$$

T is used here as the temperature of the digester in degrees Centigrade which can vary between the limits of 20° and 35° C.

Since the solid biodegradable carbon is first broken down into soluble compounds before being used as food for the active solids it is necessary to estimate the fraction of the unused food in the effluent which is in the dissolved state. From recent work by O'Rourke and McCarty[54] it is estimated that the dissolved biodegradable carbon in the effluent is about equal to the organic carbon of the remaining volatile acids. If this value is set equal to DF, then an equation shown below can be used to estimate the volatile acid in the effluent.

$$\frac{1}{TD} = \frac{K_1 \cdot DF}{CDF + DF} \tag{57}$$

CDF can be calculated from the following relationship:

$$CDF = 200\ e^{+0.12(35 - T)} \tag{58}$$

The solid and dissolved organic carbon concentrations at station 13 are then found as follows:

$$DOC_{13} = DNBC_{12} + DF \qquad (59)$$

$$SOC_{13} = SNBC_{12} + F - DF \qquad (60)$$

When TD is below some minimum value the digestion process will fail. This washout point occurs when the bacteria are removed from the digester at a greater rate than they can reproduce themselves. This minimum detention time (TDMIN) is given by the reciprocal of K_1. Since it is not desirable to even approach this value in practice a safety factor of 2.5 was used which gives the following relationship for the minimum detention time:

$$TDMIN = 2.5/K_1 \qquad (61)$$

O'Rourke & McCarty[54] have shown how the detention time at which maximum waste utilization in pounds per cubic foot of digester capacity can be calculated. At higher temperatures, this value is near the point of failure but below the TDMIN value given by equation 61. At lower temperatures, however, it may be greater than TDMIN, depending upon the raw waste concentration. Since it would never be desirable to operate at a detention time below this optimum value the optimum detention time (TDOPT) is calculated in the program after O'Rourke and McCarty as follows:

$$TDOPT = (1 - (K_2/(K_2 + DIGC_{12})^{1/2}))/K_1 \qquad (62)$$

In the program, the greater of the two values TDMIN and TDOPT is used as the minimum detention time for the digester.

Gas produced in the digester is assumed to be methane (CH_4) and carbon dioxide. Since, theoretically, COD (chemical oxygen demand) must be conserved across the digester we can state that the pounds of COD in the digester gas produced each day must equal the pounds of COD in stream 12, minus the pounds of COD in stream 13. Also, if we neglect trace gases such as ammonia, H_2S, H_2, and O_2 the COD of the digester gas is associated with the methane only because the COD of carbon dioxide is zero.

The COD of methane can be found from the following chemical equation:

$$CH_4 + 2O_2 \longrightarrow CO_2 + 2H_2O \qquad (63)$$

Thus, one pound mole or 359 cubic feet ($0^\circ C$ and one atmosphere) is equivalent to two pounds moles (64 lb) of oxygen. Thus, one pound of COD is equivalent to 5.61 cu ft of methane at STP. Using the factor of (3.5) to convert carbon to COD, we can now find the daily methane production as follows:

$$CH4CFD = (DIGC12 - DIGC13)\ (3.5)(5.61)\ (8.33)Q_{12} \qquad (64)$$

Since 359/12 or 30 cu ft of methane plus carbon dioxide gas must be produced for each pound of carbon used, the volume of carbon dioxide produced is found as the difference between the total gas volume and the volume of methane gas.

Nitrogen balances made of digester influent and effluent streams have shown[55] that all of the organic nitrogen introduced into the digester can be accounted for as either ammonia nitrogen or organic nitrogen in the effluent. Nitrogen gas found in the gas from digesters is, therefore, taken as being the result of air leaks. It is known, however, that organic nitrogen containing compounds are not destroyed at as great a rate as the non-nitrogen containing compounds. McCarty has suggested the following relationship to account for this difference in the rates of decomposition:

$$DN_{13} = DN_{12} + 0.65\ SON_{12}(DIGC12 - DIGC13)/(SOC_{12} + DOC_{12}) \quad (65)$$

This equation expresses the estimate that nitrogenous compounds are destroyed at a rate which is 65% of the rate of destruction for all organic compounds. The dissolved phosphorus in the effluent from the digester is similarly expressed as follows:

$$DP_{13} = DP_{12} + SOP_{12}(DIGC12 - DIGC13)/(SOC_{12} + DOC_{12}) \quad (66)$$

The increase in alkalinity across the digester is calculated next. This increase is due mostly to the conversion of organic nitrogen to inorganic nitrogen.

The alkalinity of stream 13 can thus be found as follows:

$$ALK13 = ALK12 + (DN13 - DN12)\ 3.57/AR \quad (67)$$

Here, 3.57 is the ratio of equivalent weight of alkalinity to weight of nitrogen, or 50/14. The value AR is the alkalinity ratio and is equal to the alkalinity increase due to nitrogen conversion divided by the total alkalinity increase during digestion. The degradation of the sodium, calcium, and magnesium soaps also results in an increase in alkalinity. The alkalinity ratio probably varies from 0.6 to 0.8. However, a value of 1.0 was used to compensate for the fact that the actual formation of inorganic nitrogen during digestion may be less than that given by equation (65).

SLUDGE ELUTRIATION

The performance of the single-stage sludge washing step is specified by the solids recovery ratio (ERR) and the concentration of solids at station 15 (TSS_{15}). These parameters, which are engineering estimates of the average performance for the process, are supplied as input to the program. The wash water ratio (WRE) is also supplied as input. The volume flow for streams 14 and 15 is then computed as follows:

$$ERR = \frac{Q_{15}\ TSS_{15}}{Q_{13}\ TSS_{13}} \tag{68}$$

$$Q_{15} = ERR \cdot Q_{13} \cdot TSS_{13}/TSS_{15} \tag{69}$$

$$Q_{17} = WRE\ Q_{13} \tag{70}$$

$$Q_{14} = Q_{13} + Q_{17} - Q_{15} \tag{71}$$

The total solids concentration of stream 14 is then found as follows:

$$TSS_{14} = Q_{13}\ (TSS_{13}\ [1 - ERR] + WRE\ TSS_5)/Q_{14} \tag{72}$$

The assumption is made here that the proportions of the various classes of solids at stations 14 and 15 equal the proportions at station 13. For example,

$$SOC_{14} = (TSS_{14}/TSS_{13})\ SOC_{13} \tag{73}$$

$$SOC_{15} = (TSS_{15}/TSS_{13})\ SOC_{13} \tag{74}$$

This assumption may or may not be adequate, but no information has been found on the way various classes of solids are distributed between the effluent streams from the elutriation step.

The concentrations of dissolved species at station 15 are found by assuming complete mixing in the elutriation tank. For example, the following material balance can be written for dissolved carbon:

$$Q_{13}\ DOC_{13} + Q_{17}\ DOC_5 = Q_{14}\ DOC_{14} + Q_{15}\ DOC_{15} \qquad (75)$$

Assuming $DOC_{14} = DOC_{15}$, we can then find the dissolved carbon at station 15 as follows:

$$DOC_{15} = \left[\frac{Q_{13}}{Q_{13} + Q_{17}}\right] DOC_{13} + \left[\frac{Q_{17}}{Q_{13} + Q_{17}}\right] DOC_5 \qquad (76)$$

All other dissolved species, including alkalinity, are computed in a similar way. It is clear that information necessary to write reliable material balances around the elutriation step is not now available. This information could be obtained from measurements of solid and dissolved species or full-sized elutriation tanks.

VACUUM FILTRATION AND SLUDGE DRYING BEDS

As suggested by McCarty,[57] the loading parameter for the vacuum filter was taken as the liquid loading rate, gph/sq ft instead of the usual solids loading rate, lb/day/sq ft. In reference 57, McCarty shows an empirical relationship between the concentration of sludge entering the vacuum filter and the moisture content of the filtered sludge. Using WP to indicate the percent moisture of the filtered sludge, the following relationship was derived to fit the empirical data:

$$WP = [illegible].0\ (TSS_{15}/10{,}000)^{-0.123} \qquad (77)$$

Letting W_w equal the pounds of water per day, and W_s equal the pounds of solids per day, the following mass balances can be written for water and solids:

$$W_{s15} = W_{s16} + W_{s18} \tag{78}$$

$$W_{w15} = W_{w16} + W_{w18} \tag{79}$$

Since

$$W_{w18} = WP(W_{w18} + W_{s18})/100 \tag{80}$$

The solids at station 18 can be expressed as follows:

$$W_{s18} = W_{w18}\ (100 - WP)/WP \tag{81}$$

However, the ratio of solids to water at station 18 can also be written as follows:

$$\frac{Q_{15}\ TSS_{15} - Q_{16}\ TSS_{16}}{1,000,000(Q_{15} - Q_{16})} = \frac{W_{s18}}{W_{w18}} \tag{82}$$

Combining equation (4) and (5) we can solve for Q_{16} as follows:

$$TEMP1 = 1,000,000\ (100 - WP)/WP \tag{83}$$

$$Q_{16} = Q_{15}(TEMP1 - TSS_{15})/(TEMP1 - TSS_{16}) \tag{84}$$

The total solids at station 18 in pounds per day is then found as follows:

$$W_{s18} = 8.33(Q_{15}TSS_{15} - Q_{16}TSS_{16}) \tag{85}$$

The relative proportions of the various species of solids at station 16 and 18 are set equal to the relative proportions at station 15. Dissolved species at station 16 are assumed equal to the dissolved species at station 15. The pounds per day of dissolved species at station 18 are found from the following mass balance equation:

$$\text{Pounds of dissolved carbon at } 18 = DOC_{15}(Q_{15} - Q_{16})8.33 \qquad (86)$$

Finally, letting VFL equal the input value of loading in gph/sq ft and letting TVF equal the fraction of each day the filter is in operation, we can find the required filter area (AVF) in sq ft as follows:

$$AVF = \frac{Q_{16}(1,000,000)}{24.0 \; VFL \; TVF} \qquad (87)$$

The amount of chemicals required to prepare the sludge for vacuum filtration was computed using the method reported by Genter[56]. The chemical used was ferric chloride. The alkalinity of the digested sludge, expressed as a percentage of the total solids (Y1) and (Y2), the ratio of volatile matter to ash in the sludge, was computed as follows:

$$Y1 = 100(ALK_{15})/TSS_{15} \qquad (88)$$

$$Y2 = VSS_{15}/SFM_{15} \qquad (89)$$

Genter shows that the total chemical demand can then be computed as follows, where FECL3 is the ferric chloride demand expressed as a percentage of the total solids:

$$FECL3 = 1.08(Y1) + 2.0(Y2) \qquad (90)$$

Letting DFECL3 equal the cost of ferric chloride per pound, the yearly cost of chemicals is then computed as follows:

$$\text{Yearly Cost, dollars} = TSS_{15}(FECL3/100)\ Q_{15}(365)(8.33)CFECL3 \qquad (91)$$

The cost of ferric chloride was taken as eight cents per pound.

Sludge drying beds were included in the program as an alternative to vacuum filters. The beds were sized using a loading factor (SBL) of 4.4 pounds of solids per sq ft of bed area per month. The capital cost of sludge drying beds was taken as $2.23 per sq ft.

SLUDGE INCINERATION

Separate relationships for capital and operating cost of incineration were not found in the literature.[59-64] Instead, the total cost of sludge incineration was given as a function of the amount of sludge incinerated per year. Letting YTONS equal the total number of tons of dry sludge solids incinerated per year, assuming the moisture content is about 70-75%, the following cost relationship was found:

$$\text{Total Cost, dollars/yr.} = 35(YTONS) - 0.000195(YTONS)^2 \qquad (92)$$

From a very limited amount of data supplied by Nichols Engineering and Research Corp. the capital cost for rotating hearth incinerators and related air pollution control equipment was estimated as follows:

$$\text{Capital Cost, dollars} = 1570.(\text{Pounds of Dry Solids/day})^{.6} \qquad (93)$$

The operating and maintenance cost was then estimated as follows:

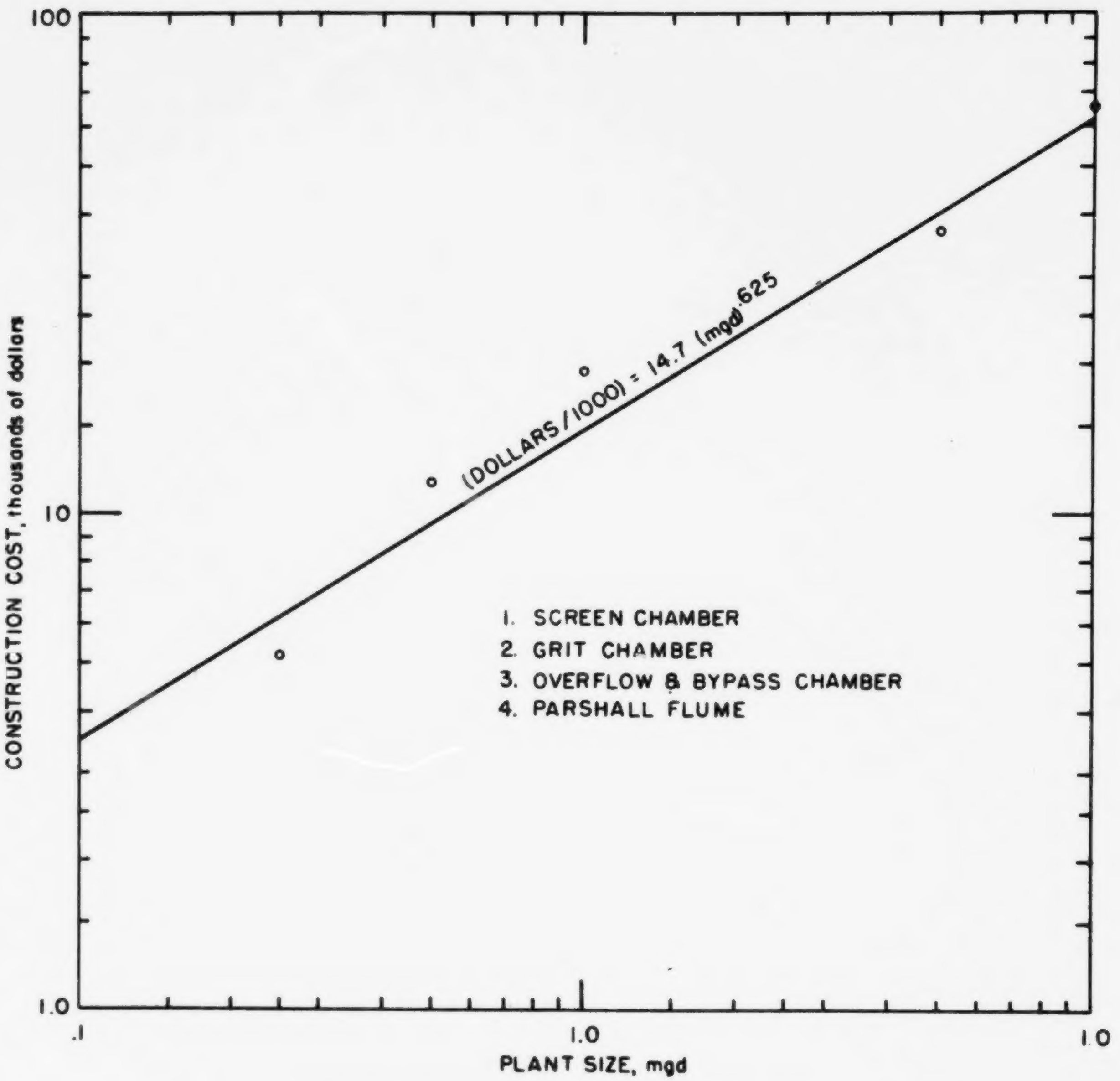
$$\text{O \& M Cost, dollars/yr.} = 16.1(YTONS) - 0.00009(YTONS)^2 \qquad (94)$$

This is recognized as a totally inadequate method for estimating incineration cost and plans exist for development of a valid model for the sludge incineration process.

CAPITAL COST FOR EQUIPMENT AND STRUCTURES

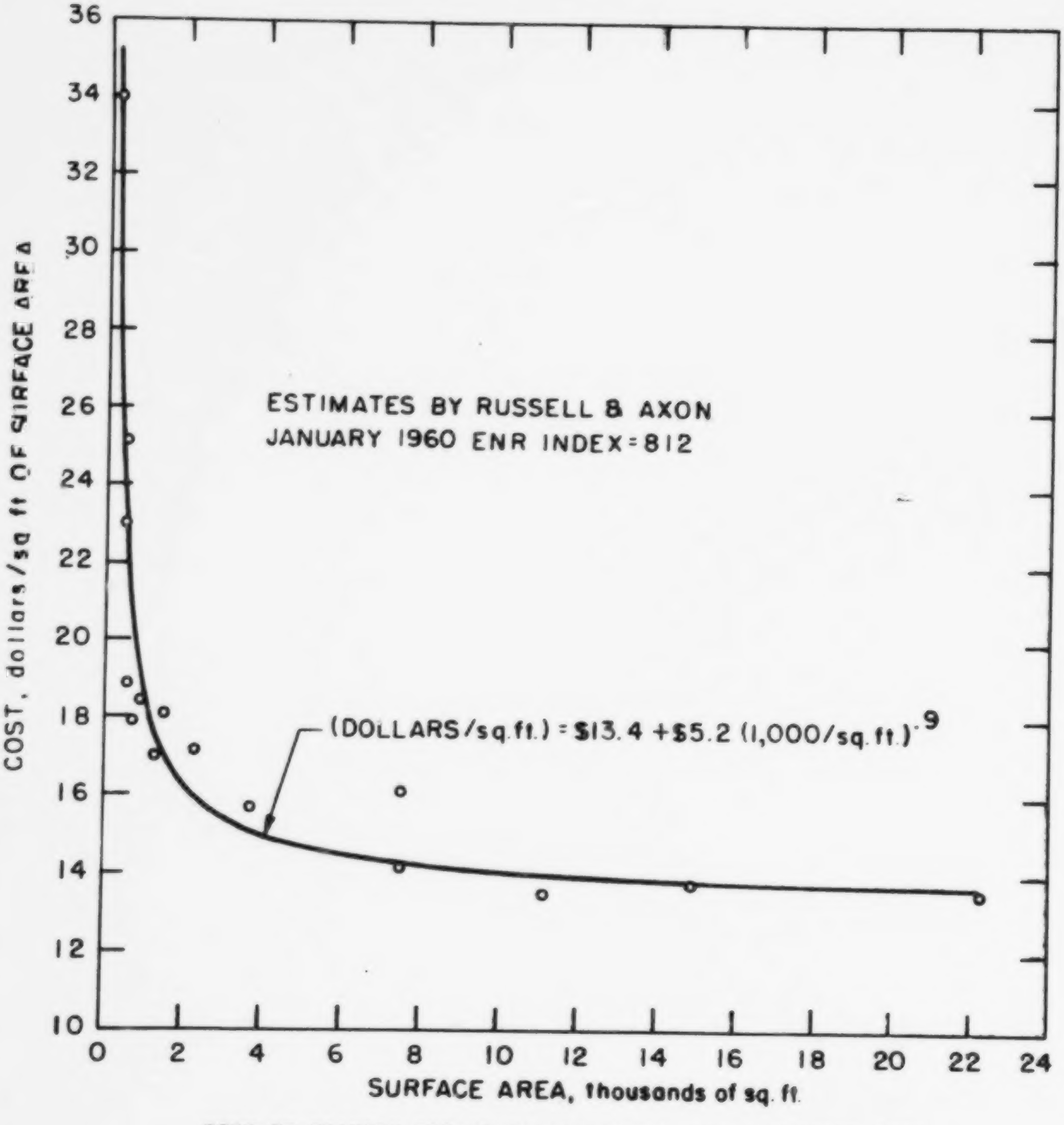
Most of the cost estimates for equipment and structures were obtained from the study conducted at Northwestern University under a United States Public Health Service research grant[65]. The actual cost estimates were made by George S. Russell of the consulting firm of Russell & Axon. Mr. Russell prepared preliminary designs for 34 individual sewage treatment plants in the range of 0.25 to 10 mgd. These plants provided preliminary treatment, primary settling, activated sludge, digestion, sludge drying beds, vacuum sludge filters, and chlorination. Conventional cost estimating techniques were used to estimate the cost of the individual unit processes. These unit process costs were then summed to find the total cost of a single plant. For this study, Russell's unit process costs were plotted versus the key design parameter for the unit process, and functional relationships were fitted to the data using graphical techniques. These relationships were used in the digital program to compute the unit process capital cost as a function of the size of the process. For example, the cost of tanks is given in terms of tank volume or in terms of surface area. Cost of pumps or blowers is given in terms of the design point capacity in cubic feet per minute or gallons per minute. The derived relationships are shown with the estimated cost in Figures 4 through 15. **These costs are** keyed to an ENR index of 812, which applies to January 1960. A capital cost index (CCI) is supplied to the program as input to adjust capital cost to current values.

The Ten-State Standards[67] were used by Russell as a basis of design in most cases. The grit chamber was designed to have a flow-through velocity of 1.0 ft/sec at maximum flow, which was taken as 3.25 times design flow. Primary sedimentation was based on 1, 2, and 3 hours of detention with a check to made sure that the overflow rate is below 800 gpd/sq ft. The



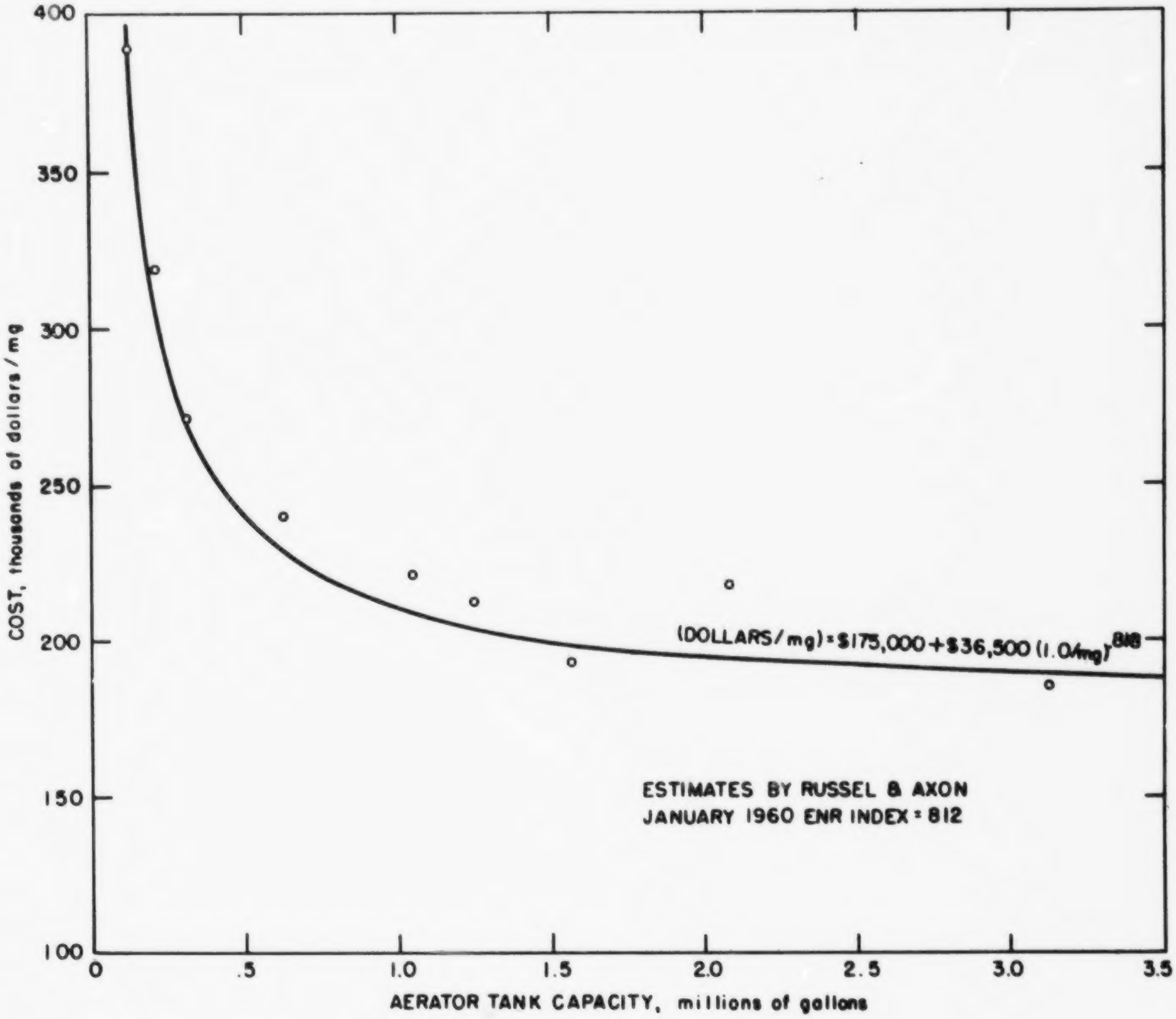
COST OF PRELIMINARY TREATMENT versus PLANT SIZE mgd

Figure 4



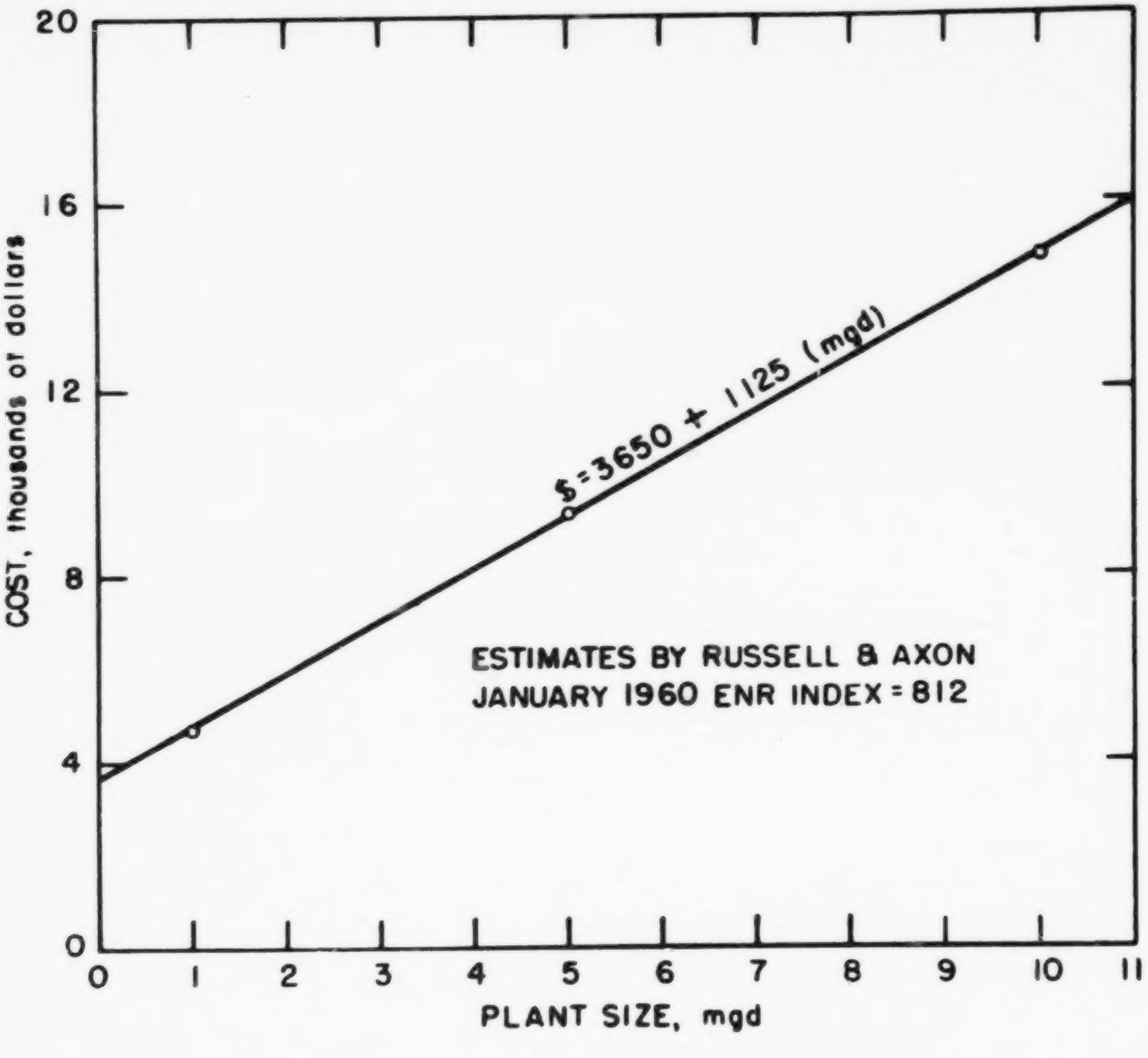
COST OF PRIMARY SEDIMENTATION TANKS versus SURFACE AREA

Figure 5



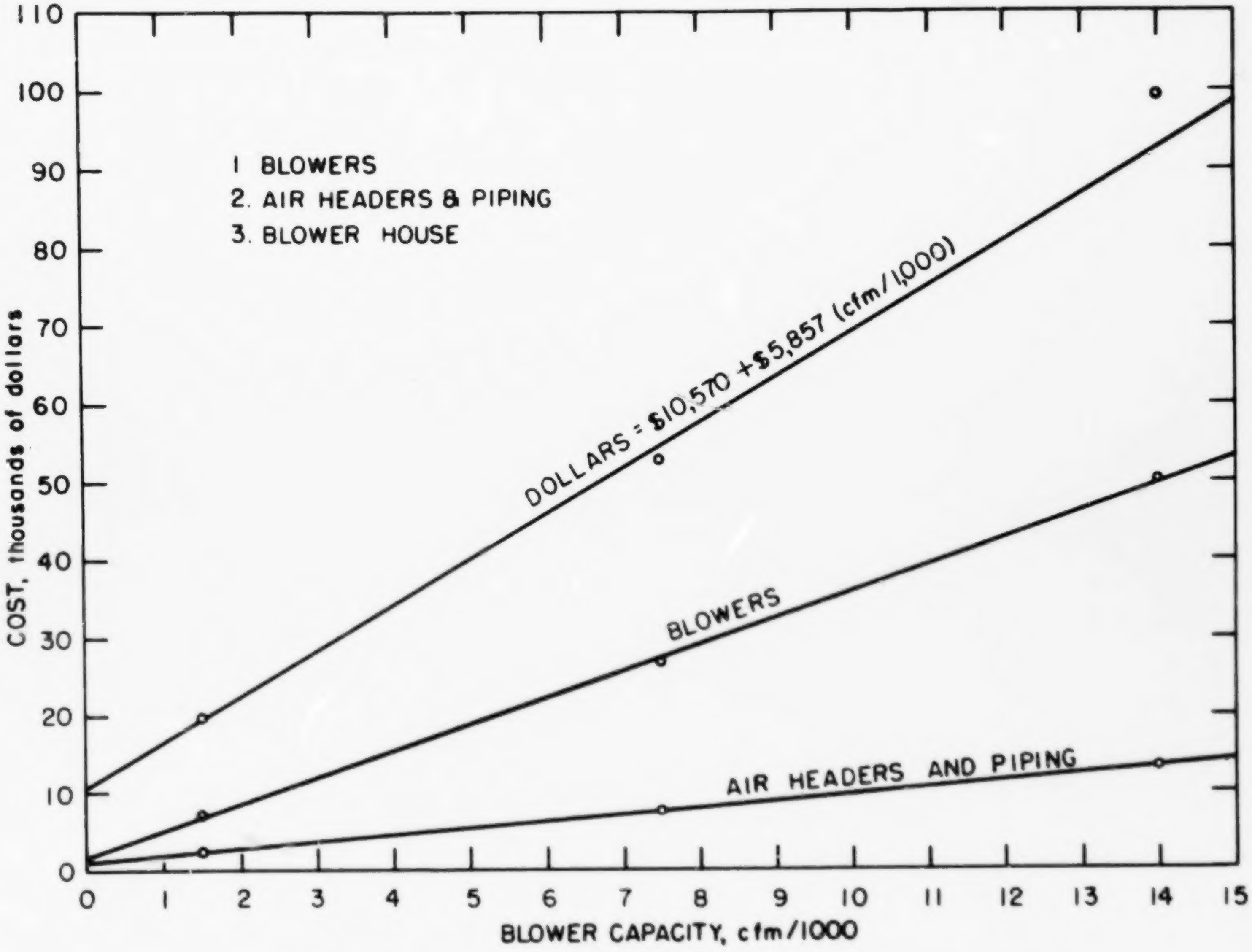
COST OF AERATORS AND PIPE GALLERY

Figure E



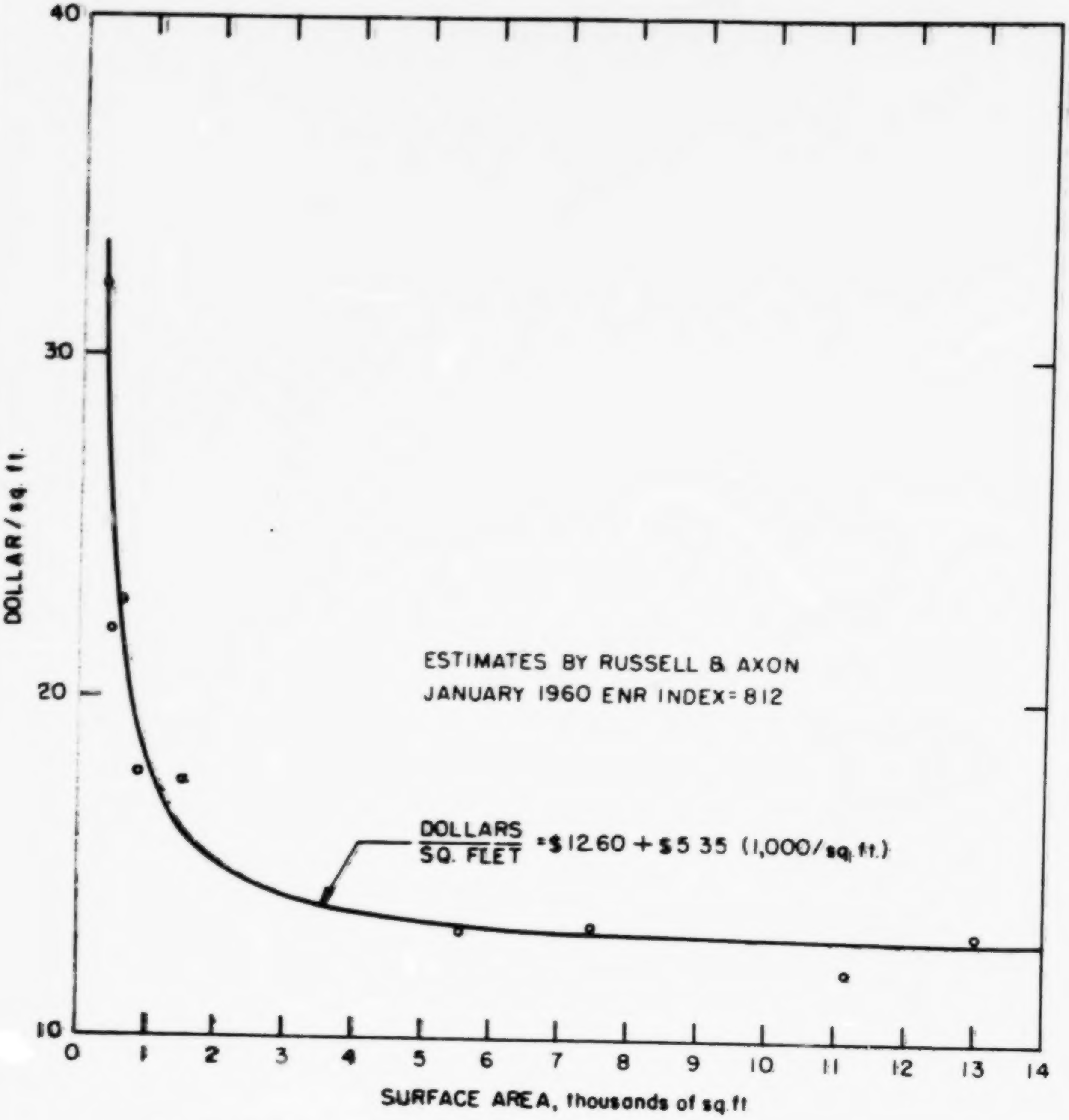
COST OF ACTIVATED SLUDGE RETURN PUMPS

Figure 7



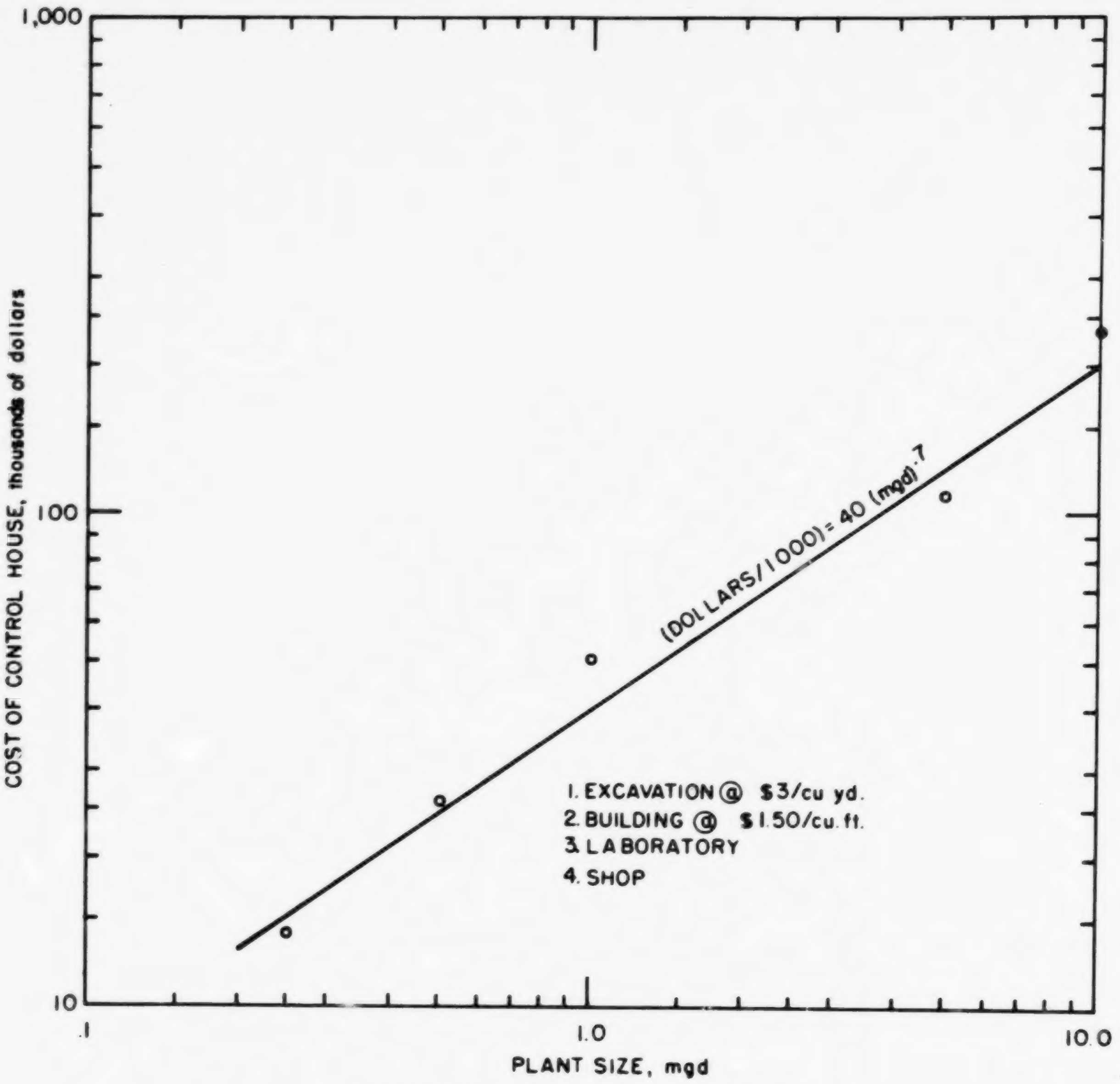
CONSTRUCTION COST FOR BLOWER INSTALLATION versus BLOWER CAPACITY

Figure 8



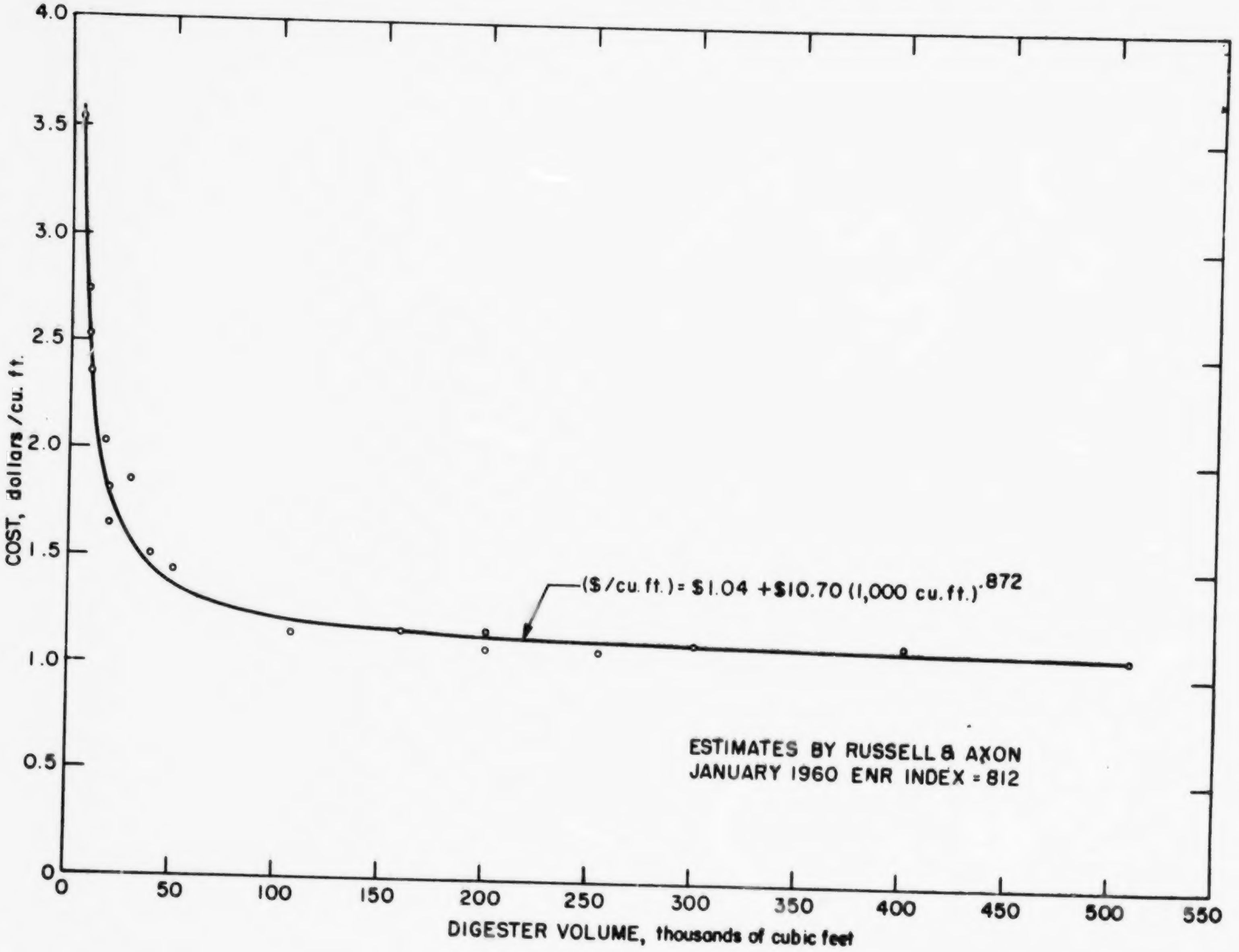
COST OF FINAL SETTLING TANKS INCLUDING PIPES, VALVES AND FITTINGS - SLUDGE WASTE PUMP SLUDGE COLLECTOR

Figure 9



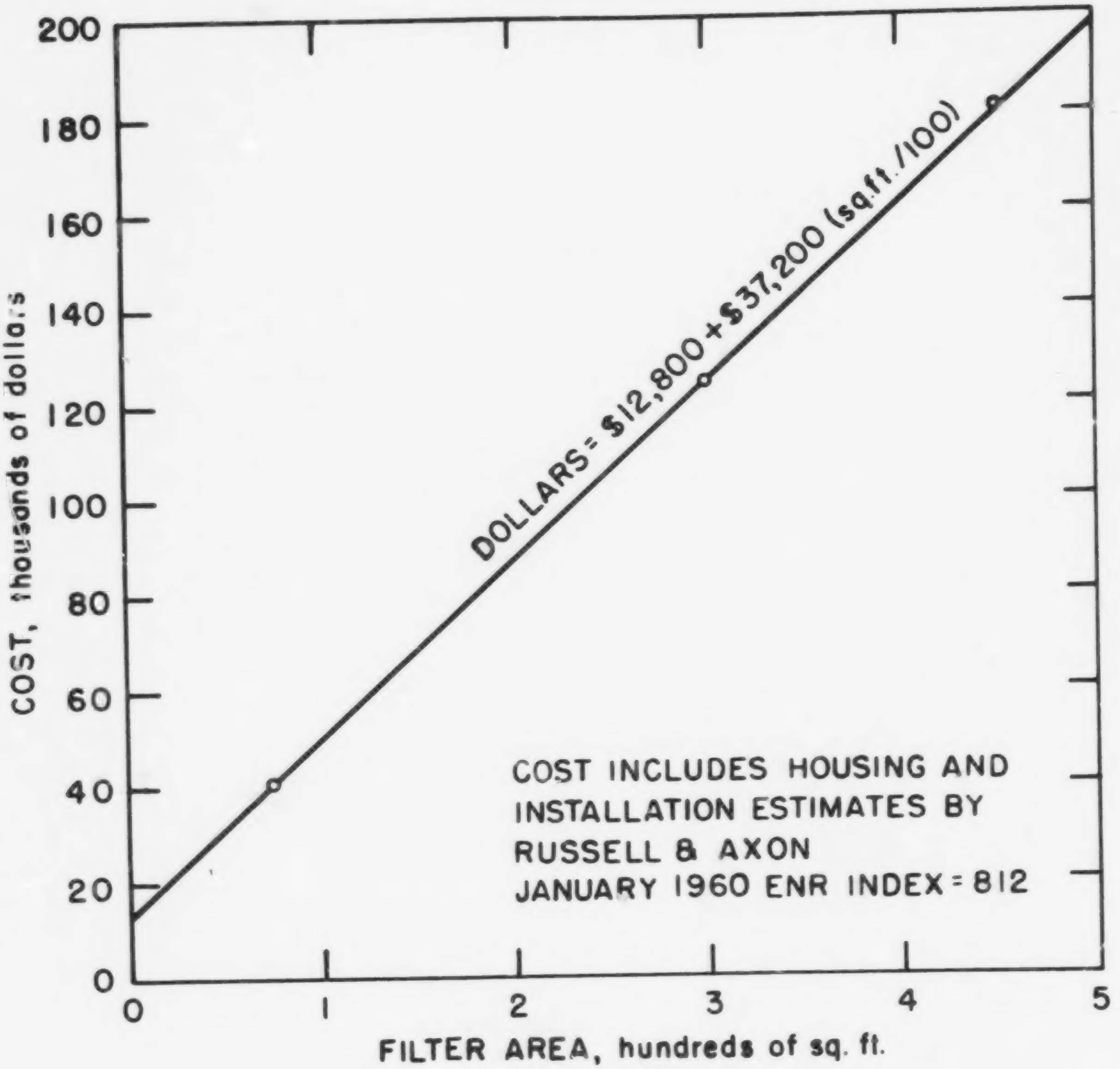
COST OF CONTROL HOUSE versus PLANT SIZE

Figure 10



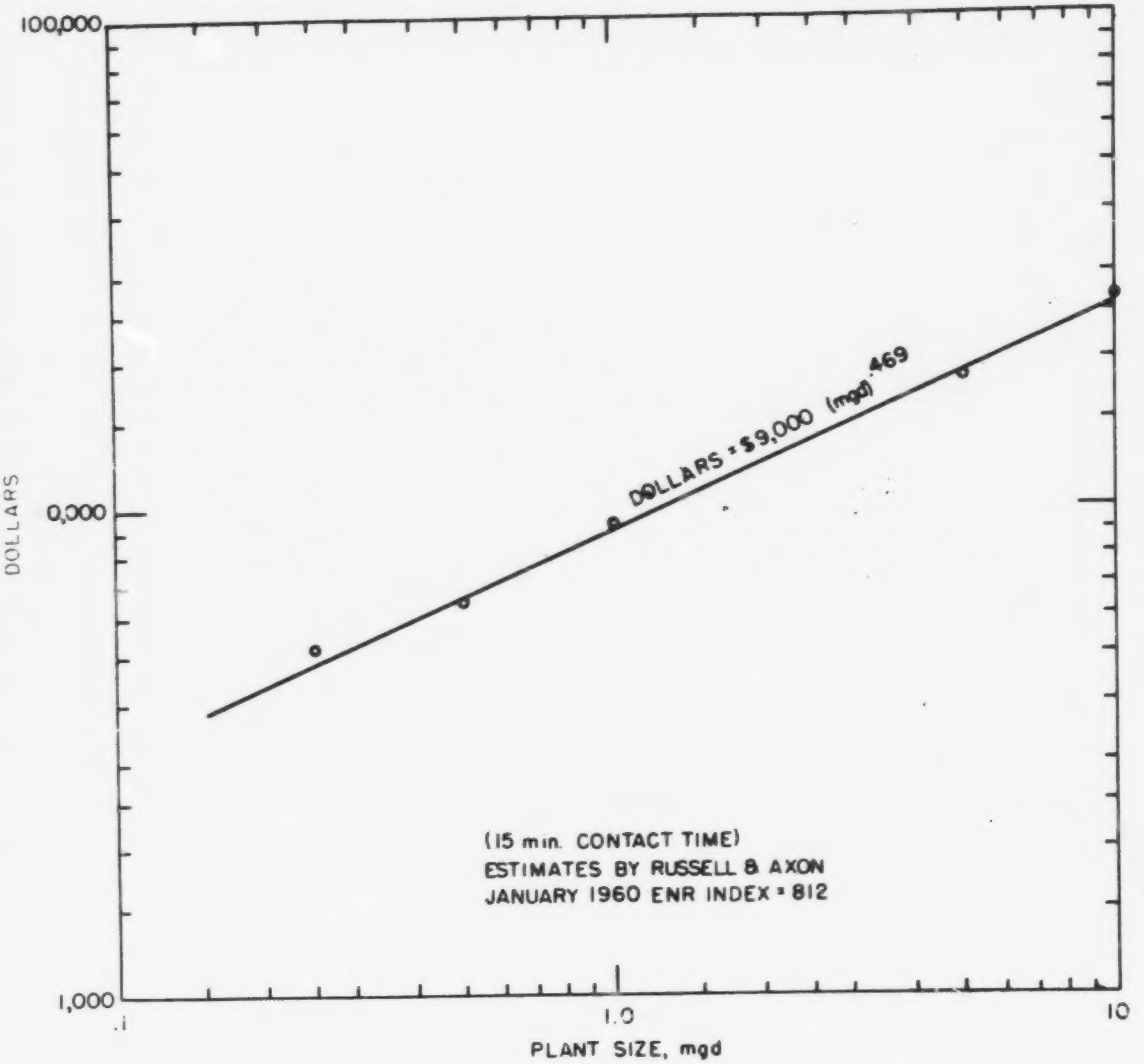
COST OF DIGESTERS versus DIGESTER VOLUME

Figure 11



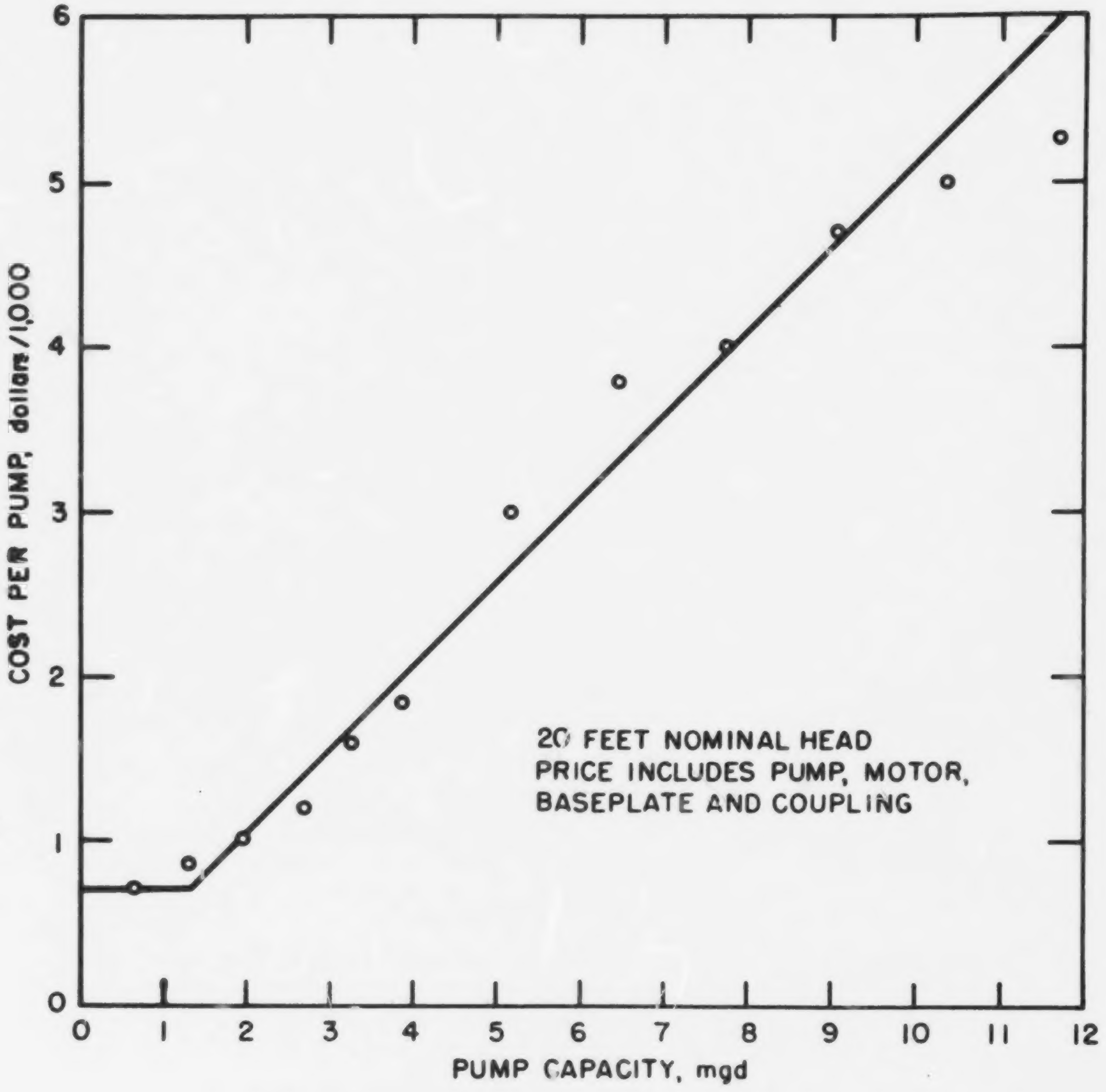
COST OF VACUUM SLUDGE FILTERS versus AREA OF FILTER

Figure 12



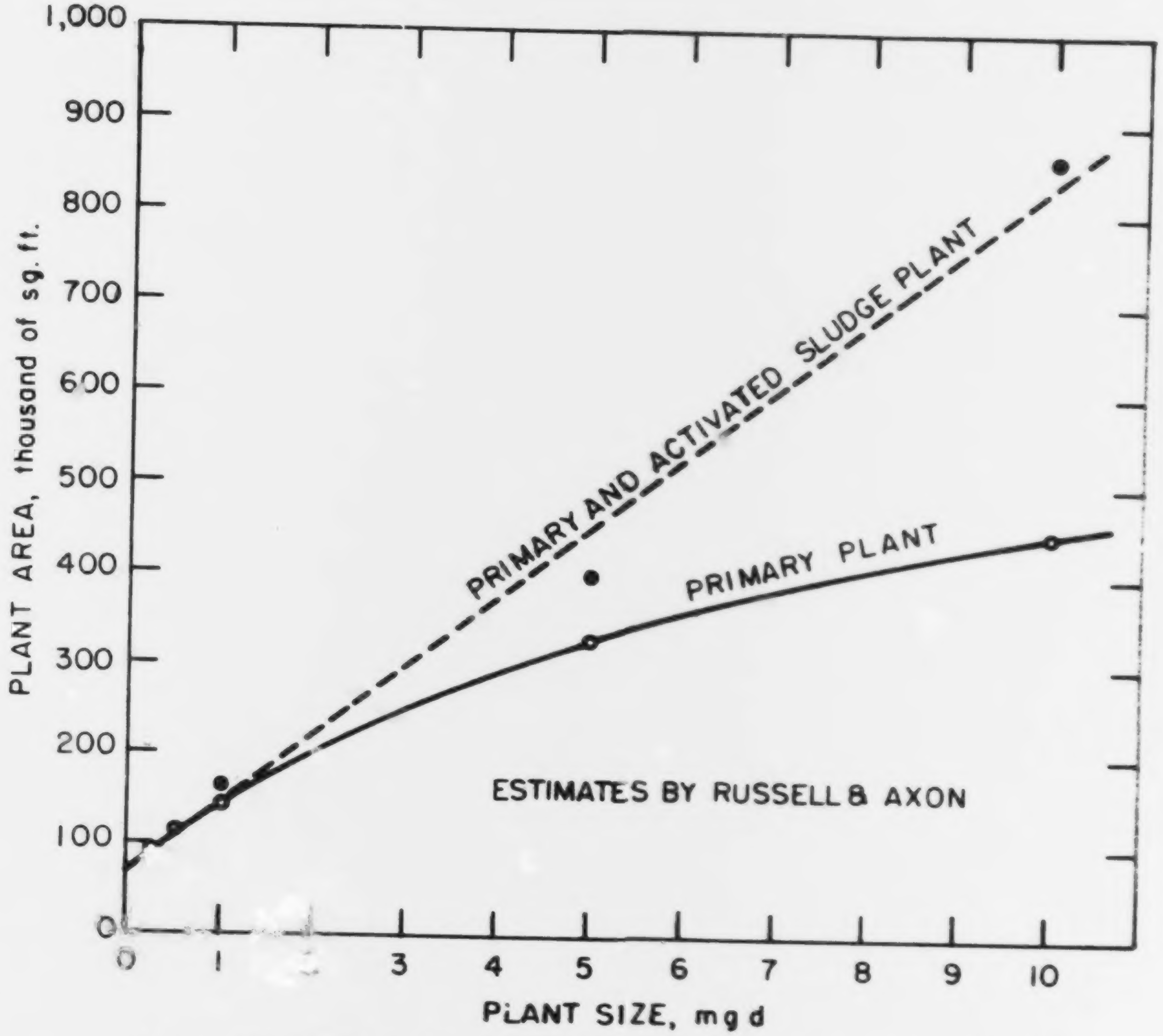
COST OF CHLORINE CONTACT TANK AND CHLORINATOR versus PLANT SIZE

Figure 13



COST OF INFLUENT PUMPS versus FLOW CAPACITY (mgd)

Figure 14



LAND REQUIREMENTS FOR PRIMARY AND PRIMARY WITH ACTIVATED SLUDGE PLANTS versus PLANT SIZE

Figure 15

depth of primary sedimentation tanks was usually 7.5 ft with several designs calling for a 6.5-ft depth. Plants, including primary sedimentation, anaerobic digestion, sludge beds and chlorination, were designed for flows of 0.25, 0.50, 1.0, 5.0, and 10.0 mgd. Three sizes of activated sludge plants were designed: 1.0, 5.0, and 10.0 mgd. Digesters for primary plants were based on 2 cu ft/capita. One hundred gallons of sewage per day per capita were assumed. Aeration tanks for the activated sludge process were designed for 3, 5, and 7.5 hours of detention (assuming no return flow). The aerators were all 15 ft deep. Digesters for activated sludge plants were based on 5 cu ft per capita. Sludge beds provided 1.75 sq ft per capita. The activated sludge return pumps were sized to return 50% of the design flow. Air blowers were specified to supply 2 std cu ft of air per gallon of sewage. Aero-degritters were given as an alternative to grit chambers. but the air used by the degritters was a negligible fraction of the supply. The final settling tanks for the 1.0 mgd-plant were based on 800 gpd/sq ft of design flow. The 5 and 10-mgd final settling tanks were based on 1200 gpd/sq ft, but a safety factor of 1.25 made the effective design overflow rate equal to 960 gpd/sq ft. The depth of the final settling tanks varied from 5 to 8.5 ft. Vacuum sludge filters were based on 6 lb sludge solids per hour per sq ft. Two hundred mg/l suspended solids and 6 hours/day operating time were assumed. Land requirements were based on a plan veiw sketch of the plant. Chlorine contact tanks were designed to give 15-minutes contact time.

An itemized list of the factors considered in making the construction cost estimates is given below:

I. PRELIMINARY TREATMENT

1. Screen Chamber
    - a. Excavation @ $3.00 per cu yd
    - b. Reinforced concrete @ $125 per cu yd
    - c. Misc. iron @ $0.20 per lb
    - d. Comminutor
2. Aero-Degritter
    - a. Excavation @ $3.00 per cu yd
    - b. Reinforced concrete @ $85.00 per cu yd
    - c. Air lift tubes, etc.
    - d. Misc. iron and steel @ $0.20 per lb
3. Overflow and Bypass Chamber
    - a. Excavation @ $3.00 per cu yd
    - b. Reinforced concrete @ $125 per cu yd
    - c. Sluice Gate
    - d. Misc. iron and steel @ $0.20 per lb
4. Parshall Flume
    - a. Excavation @ $3.00 per cu yd
    - b. Reinforced concrete @ $125 per cu yd
    - c. Small piping to control house
    - d. Indicating and integrating meter

II. PRIMARY SEDIMENTATION

- a. Excavation @ $3.00 per cu yd
- b. Reinforced concrete @ $85.00 per cu yd
- c. Misc. iron and steel @ $0.20 per lb
- d. Pipe, valves, and fittings
- e. Sludge pumps
- f. Motor driven collector and skimmer
- g. Scum collector pipes

III. DIGESTERS

- a. Excavation @ $3.00 per cu yd
- b. Reinforced concrete @ $85.00 per cu yd
- c. Digester house @ $1.50 per cu ft
- d. Supernatant, etc., @ $8.00 per linear ft
- e. Small piping and valves including heating coils

- f. Sludge circulating pump
- g. Floating digester cover
- h. Pipe to sludge beds @ $8.00 per linear ft

IV. SLUDGE BEDS

- a. Excavation @ $3.00 per cu yd
- b. Reinforced concrete @ $85.00 per cu yd
- c. Sand and gravel @ $3.00 per cu yd
- d. Sludge piping @ $5.00 per linear ft
- e. Underdrains @ $3.00 per linear ft
- f. Misc. iron and steel @ $0.20 per lb

V. SLUDGE HEATING

- a. Hot water circulating pump
- b. Heat exchanger
- c. Hot water and gas piping

VI. SECONDARY ACTIVATED SLUDGE PROCESS

1. Aerators and pipe gallery
    - a. Excavation @ $3.00 per cu yd
    - b. Reinforced concrete @ $90.00 per cu yd
    - c. Air diffusers
    - d. Pipes, valves, and fittings
    - e. Misc. iron and steel @ $0.20 per lb

2. Return Activated Sludge Pumps
    - a. Pumps
    - b. Controls and piping

3. Blowers and Blower House
    - a. Blowers
    - b. Air headers and piping
    - c. Blower house @ $1.50 per cu ft

VII. FINAL SEDIMENTATION TANKS (800-1200 gpd/ft^2)

- a. Excavation @ $3.00 per cu yd
- b. Reinforced concrete @ $85.00 per cu yd
- c. Misc. iron and steel
- d. Pipes, valves, and fittings
- e. Waste sludge pump
- f. Motor driven collector

VIII. CHLORINATION (15-minute contact time)

- a. Excavation @ \$3.00 per cu yd
- b. Reinforced concrete @ \$85.00 per cu yd
- c. Misc. iron and steel @ \$0.20 per lb
- d. Chlorinator

IX. VACUUM SLUDGE FILTRATION (Substitute for sludge beds)

- a. Filters
- b. Additional control house volume

The capital cost relationships derived from Russell's plant estimates are summarized below:

UNIT PROCESS	CAPITAL COST, dollars
Preliminary Treatment	$14{,}700(Q_1)^{0.625}$
Primary Sedimentation	$13{,}400(APS) + 5200(APS)^{0.1}$
Aerator	$175{,}000(VAER) + 36{,}500(VAER)^{0.182}$
Air Blowers	$10{,}750 + 5{,}857(BSIZE)$
Final Settler	$12{,}600(AFS) + 5350/(AFS)^{0.126}$
Sludge Return Pumps	$3650 + 1125(Q_6)$
Control House	$40{,}000(Q_1)^{0.7}$
Digesters	$5{,}000 + 1080(VDIG) + 10{,}700(VDIG)^{0.128}$
Vacuum Filters	$12{,}800 + 372(AVF)$
Sludge Drying Beds	$2.23(ASB)$
Chlorine Contact Tank	$9000(Q_5)^{0.469}$
Plant Site	$4400(Q_{20})^{0.875}$

Russell's work did not include the capital cost for sludge thickeners or sludge elutriation tanks. This information was obtained by fitting a number of individual points obtained from the personal file of Mr. C. Swanson of FWPCA which were based on an ENR index of 850. The following relationships were found after correcting the data to an ENR index of 812 to make the data consistent with Russell's work:

UNIT PROCESS	CAPITAL COST, dollars
Sludge Thickeners	(18.1 + 8.46/exp(ATHM/13375))ATHM
Sludge Elutriation Tanks	(18.1 + 49.5/exp(AE/6000))AE

The volumes of settlers, aerator, and digester computed by the program are based on the capacity actually needed for normal operation. No excess tank capacity is included to allow for periodic cleaning and maintenance. This is handled in the program by providing as input an excess capacity factor (ECF) for each process which multiplies the computed tank capacity values. A list of the ECF values used in the results shown in this report are given below:

UNIT PROCESS	EXCESS CAPACITY FACTOR
Preliminary Treatment	1.0
Primary Settling	2.0
Aerator	1.2
Air Blowers	1.5
Final Settler	2.0
Sludge Return Pumps	2.0
Control House	1.0
Thickener	1.5
Digesters	2.0
Sludge Elutriation	1.5
Vacuum Filter	1.0
Incinerator	1.0
Sludge Drying Beds	1.0
Chlorination	1.0
Plant Site Preparation	1.0

The capital cost of each process is based on the required capacity including the excess capacity needed for cleaning and maintenance. The total capital cost for the plant (TOTCC) is found by adding the capital costs of the individual processes. Ten percent of the total capital cost (TOTCC) is added for contractor's profit. Fifteen percent of the total capital cost is added for contingencies and omissions. The cost of engineering expressed as a percentage of the total capital cost is computed from the following relationship:

$$CENG = 8.0(1000000/TOTCC)^{0.146}$$

OPERATING AND MAINTENANCE COST

Operating and maintenance cost estimates were taken from a desk-top study made by Swanson.[69] Swanson studied annual treatment plant reports from 38 individual plants and expressed the annual operating and maintenance cost as a function of the plant flow, mgd, and as a function of the key design parameter for the individual unit processes. The following definitions taken from Swanson's memorandum show clearly what items are included in the operating and maintenance cost:

Cost Definitions

1. Direct Costs - Costs directly attributable to the unit process, i.e., operating labor and supplies, maintenance labor and materials, indirect labor costs (retirement, insurance, etc.), and power. Power is considered separately for raw sewage pumping and activated sludge processes.
2. Plant Overhead - Costs of common services and supervision rendered at the plant site to the unit processes. Included are supervision, general plant vehicles, laboratory, stores, maintenance of roads and grounds, telephone, plant administrative building, etc.

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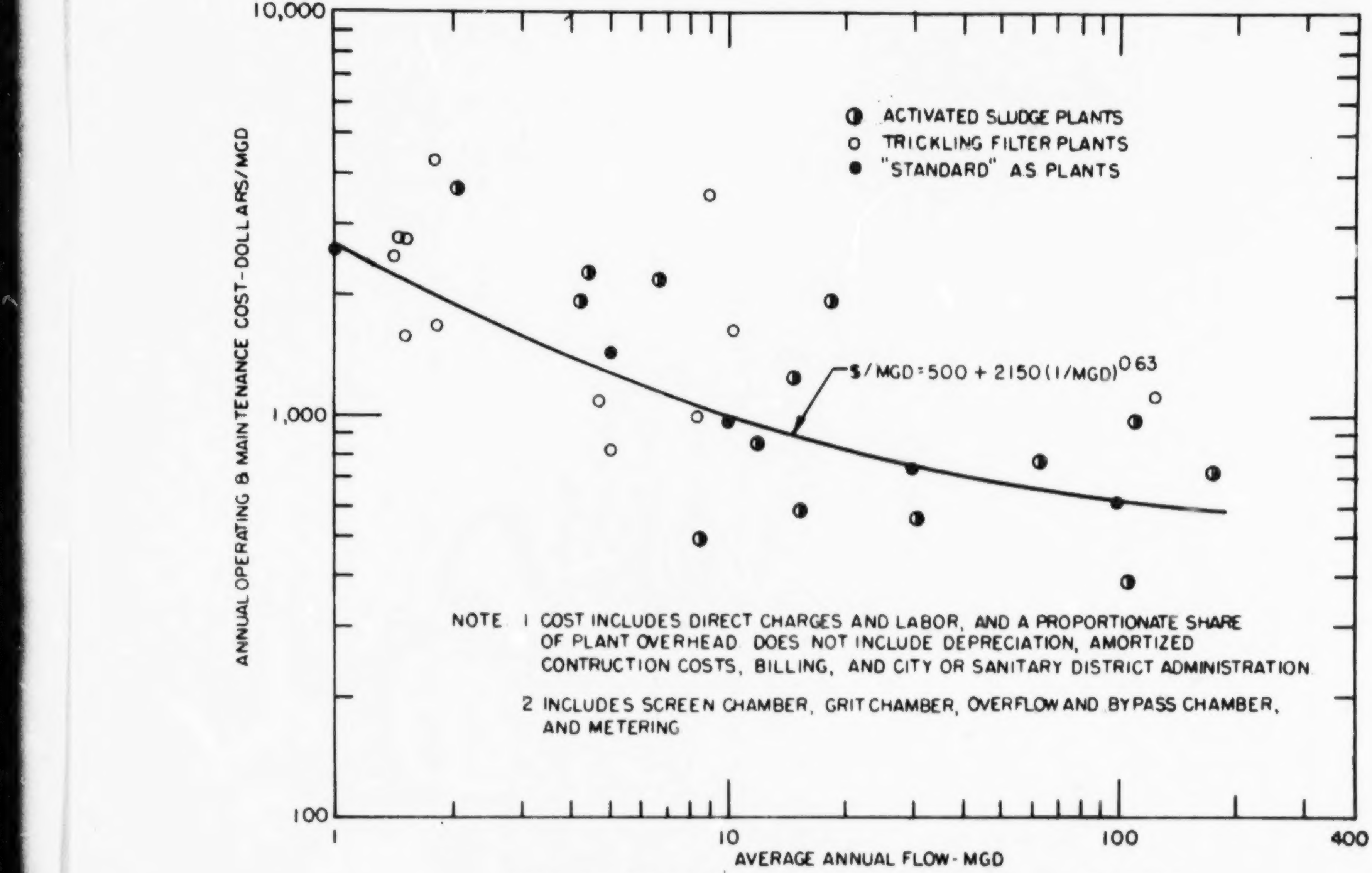
3. Administrative Costs - A share of city or sanitary district administration and billing attributable to the treatment plant. Included are general supervision, personnel services, insurance, billing and auditing, legal and financial, and engineering. Administration costs are considered as a lump sum and are not distributed to the unit processes.

4. Operating and Maintenance Cost - Sum of direct costs and plant overhead. Plant overhead is distributed to the unit processes on the basis of percent of the total direct costs attributable to the unit process.

The following relationships, taken from Swanson's work, were used to compute operating and maintenance cost in the computer program.

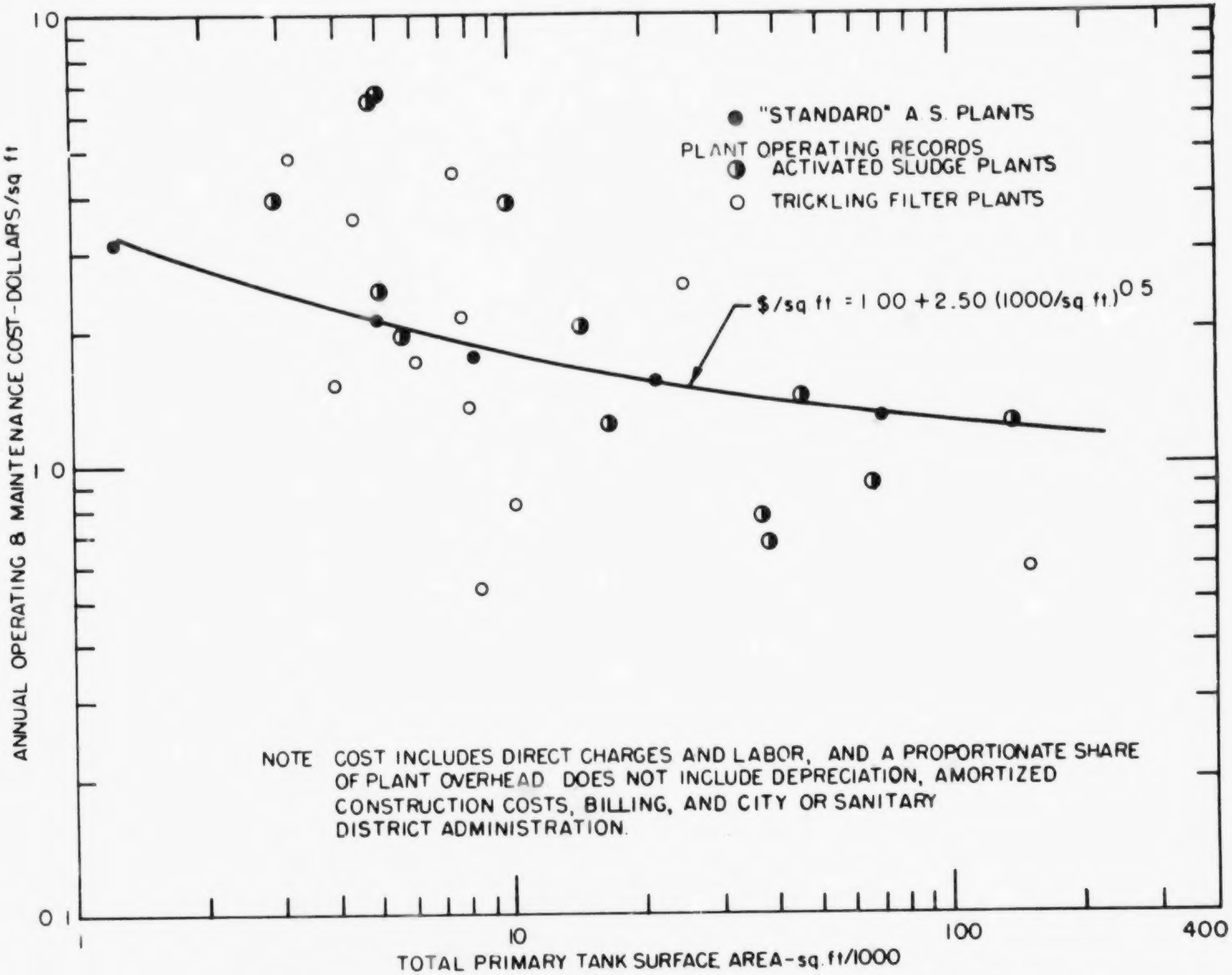
UNIT PROCESS	COST RELATIONSHIP
Preliminary Treatment	Dollars/yr = $500(Q_1) + 2150(Q_1)^{0.37}$
Primary Settling	Dollars/yr = $1000(APS) + 2500(APS)^{0.5}$
Aerator, Final Settler Blowers - excluding power	Dollars/yr = $10{,}000(VAER) + 14{,}500(VAER)^{0.63}$
Digesters	Dollars/yr = $48(VDIG) + 540(VDIG)^{0.44}$
Digesters and Sludge Drying Beds	Dollars/yr = $80(VDIG) + 900(VDIG)^{0.44}$
Vacuum Filter - excluding chemicals	Dollars/yr = $1500(Q_1) + 6450(Q_1)^{0.37}$

The cost of electrical power used by the blowers was taken as one cent per kilowatt hour. The cost of ferric chloride for sludge conditioning was taken as eight cents per pound. No credit was given for sludge gas produced and no charge was made for heating the digester. Three representative charts from Swanson's desk-top study are shown in Figures 16, 17, and 18.



PRELIMINARY PROCESS OPERATING AND MAINTENANCE COSTS versus PLANT SIZE

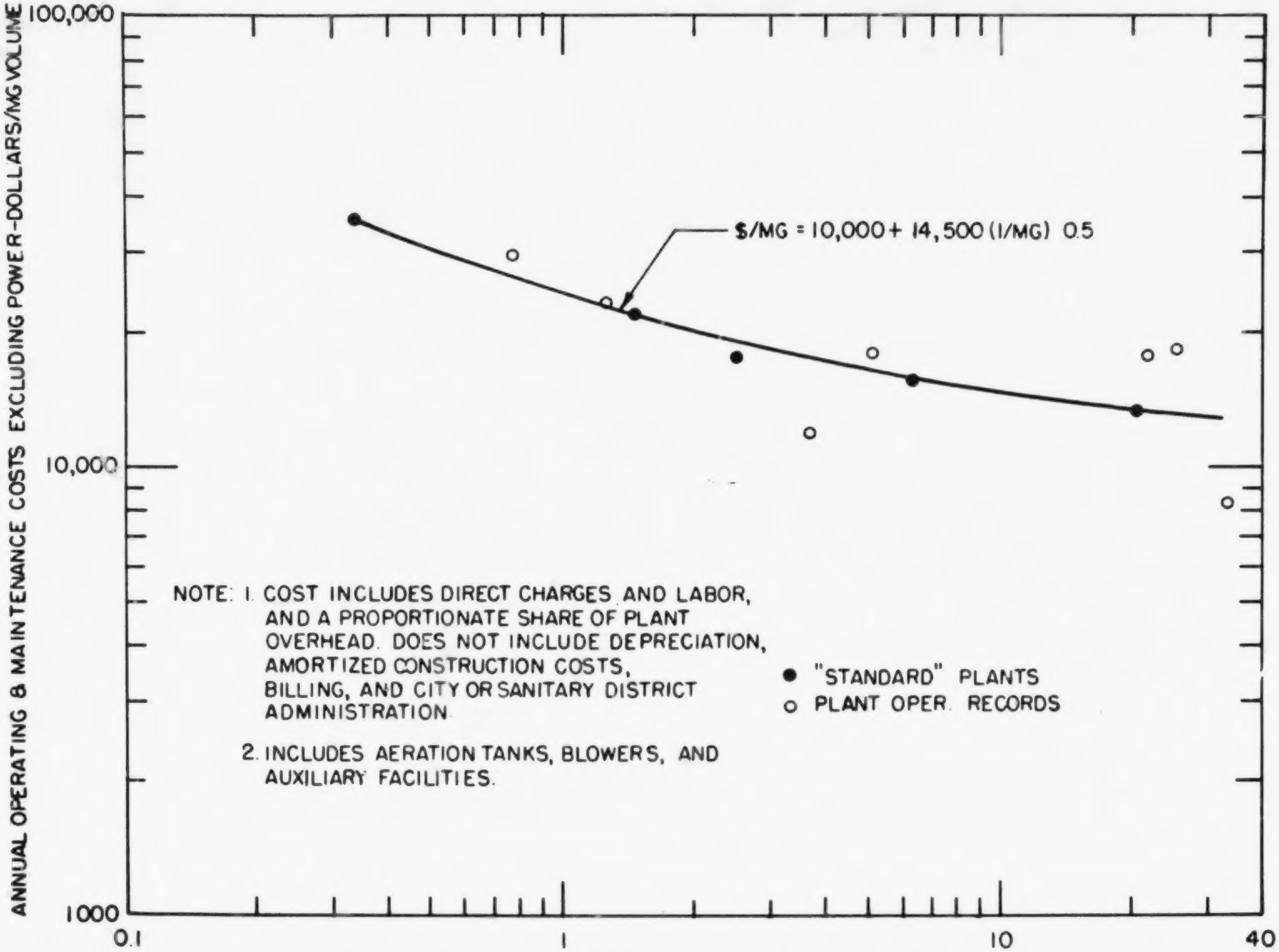
Figure 16



NOTE: COST INCLUDES DIRECT CHARGES AND LABOR, AND A PROPORTIONATE SHARE OF PLANT OVERHEAD. DOES NOT INCLUDE DEPRECIATION, AMORTIZED CONSTRUCTION COSTS, BILLING, AND CITY OR SANITARY DISTRICT ADMINISTRATION.

PRIMARY PROCESS OPERATING AND MAINTENANCE COSTS versus TOTAL TANK SURFACE AREA

Figure 17



NOTE: 1. COST INCLUDES DIRECT CHARGES AND LABOR, AND A PROPORTIONATE SHARE OF PLANT OVERHEAD. DOES NOT INCLUDE DEPRECIATION, AMORTIZED CONSTRUCTION COSTS, BILLING, AND CITY OR SANITARY DISTRICT ADMINISTRATION.

2. INCLUDES AERATION TANKS, BLOWERS, AND AUXILIARY FACILITIES.

ACTIVATED SLUDGE PROCESS OPERATING AND MAINTENANCE COSTS EXCLUDING POWER versus TOTAL TANK VOLUME

Figure 18

RESULTS OF COMPUTATION

Most of the computations made using the program have been for the purpose of relating cost to performance. As the program now stands, it is possible with a fair degree of reliability to compute the capital cost, amortization or debt service cost, and operating and maintenance cost of the entire plant or of individual processes or groups of processes as a function of the following basic independent design parameters:

1. Plant Design Capacity, Volume Flow at Station 20, mgd
2. Influent Stream Vector for Stream 20
3. Fraction of Suspended Solids Removed in the Primary Settler, FRPS
4. Mixed Liquor Suspended Solids Held in Aerator, MLSS, mg/1
5. Demand BOD of Plant Effluent at Station No. 5, mg/1
6. Detention Time for the Digester, TD, days

For convenience, the processes included in the plant have been grouped as follows:

I. PRIMARY GROUP
   1. Preliminary Treatment
   2. Primary Settler
   3. Control House
   4. Site Development

II. SECONDARY GROUP
   1. Aerator
   2. Air Blowers
   3. Sludge Return Pumps
   4. Final Settler

III. SLUDGE HANDLING GROUP
   1. Sludge Thickener
   2. Digester
   3. Sludge Elutriation
   4. Vacuum Filtration
   5. Incineration

IV. CHLORINATION

In the digital program the unit process to which a cost is referred is identified by a subscript. For example, CCOST(1) refers to the capital cost of the preliminary treatment and COSTO(2) refers to the operating and maintenance cost for the primary settler. The list of unit processes and the corresponding subscript is shown below:

UNIT PROCESS	SUBSCRIPT
Preliminary Treatment	1
Primary Settling	2
Aerator	3
Air Blowers	4
Final Settler	5
Sludge Return Pumps	6
Control House	7
Thickener	8
Digesters	9
Sludge Elutriation	10
Vacuum Filter	11
Incinerator	12
Sludge Drying Beds	13
Chlorination	14
Plant Site	15

A list of input values normally used in making computations with the digital computer program are shown in Table I. The concentrations of contaminants assumed for raw sewage, represented by stream 20, are given on page 9. A limited number of computations were made with all raw sewage concentrations shown on page 9 increased by a factor of (1.5) representing concentrated sewage. These computed results are shown in Table II.

TABLE I

INPUT VALUES HELD FIXED IN SYSTEM COMPUTATION

DEGC = 20°C

URPS = 400

URSS = 3.0

CAER20 = 1.0

AEFF20 = 0.050

DO = 1.0 mg/l

CKWH = 0.01 dollar

CCI = 1.13

AF = 0.06744 (25 yr-4 1/2%)

GSS = 2000 gpd/sq ft

TRR = 0.95

GTH = 750 gpd/sq ft

TSS_{12} = 6%

CLAND = 0.02

CFECL3 = 0.08 $/lb

GSTH = 9.0 lb/day/sq ft

ERR = 0.76

TSS_{15} = 6%

WRE = 3.0

GE = 800 gpd/sq ft

GES = 9.0 lb/day/sq ft

TDIG = 33°C

TD = 15 days

TSS_{16} = 200 mg/l

VFL = 4.9 gph/sq ft

TVF = 0.238

FRPS = 0.50

CCL2 = 0.08 $/lb

CTRP = 0.10

GTGO = 0.15

DCL2 = 8.0 mg/l

Table II

COMPUTED TOTAL TREATMENT COST FOR ACTIVATED SLUDGE PLANTS IN CENTS/1000 GAL.
USING
CONCENTRATED SEWAGE (300 mg/1 BOD) AS INFLUENT

Plant Size	Effluent BOD, mg/1	Primary Group	Secondary Group	Sludge Handling Group	Chlorination	Total
1 mgd	13	4.40	8.14	10.99	.79	24.3
	20	4.41	6.31	10.91	.79	22.4
	25	4.41	5.54	10.84	.79	21.6
	30	4.42	4.85	10.90	.79	21.0
3 mgd	13	3.16	6.23	8.05	.67	18.1
	20	3.17	4.66	7.99	.67	16.5
	25	3.17	4.00	7.93	.67	15.8
	30	3.18	3.40	7.99	.67	15.3
10 mgd	13	2.38	5.24	6.30	.61	14.5
	20	2.38	3.83	6.25	.61	13.1
	25	2.38	3.24	6.20	.61	12.4
	30	2.39	2.69	6.25	.61	12.0
30 mgd	13	1.95	4.80	5.30	.57	12.6
	20	1.95	3.46	5.25	.57	11.2
	25	1.95	2.90	5.22	.57	10.6
	30	1.96	2.39	5.26	.57	10.2
100 mgd	13	1.66	4.54	4.60	.55	11.4
	20	1.66	3.25	4.56	.55	10.0
	25	1.66	2.71	4.53	.55	9.5
	30	1.67	2.21	4.58	.55	9.0

NOTE: Fraction of Solids Removed in Primary Settler = 50%
Mixed Liquor Suspended Solids in Aerator = 2000 mg/1

Figure 19 was taken from Reference No. 70 and is believed to be a good average of plant costs reported in the literature. Corresponding costs as computed by the program are shown in Figure 20. In computing the cost values shown in Figure 20 it was assumed that the primary settler removes 50% of the suspended solids and 35% of the BOD. All assumptions relating to the sludge handling group of processes are shown in Table I. It will be noted that the cost computed by the program is generally higher than that shown in Figure 19. Reasons for this are not fully understood.

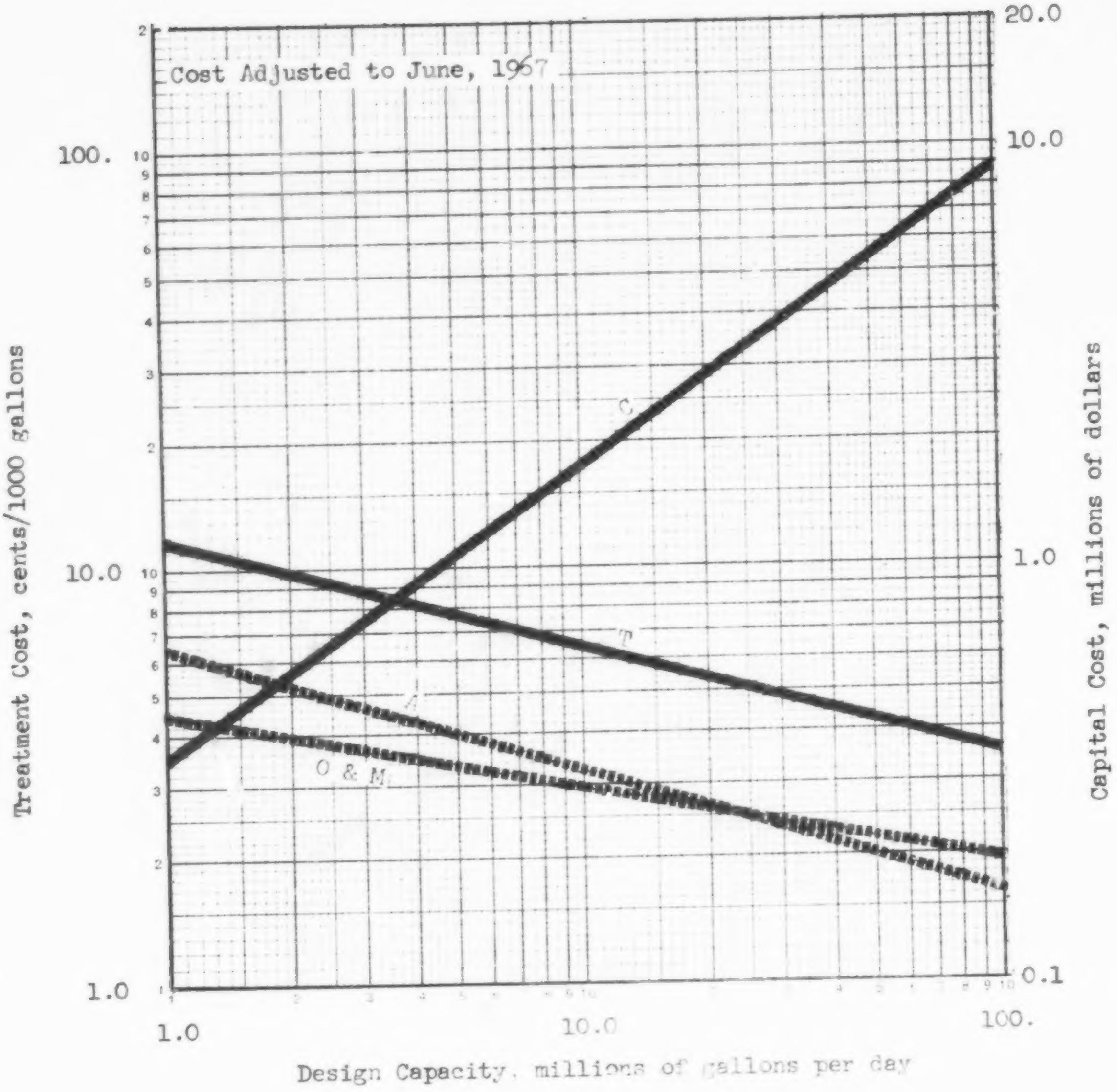
Figure 21 which was also taken from Reference No.70 shows the average cost of activated sludge plants as reported in the literature. Corresponding costs as computed by the program are shown in Figure 22. Here the agreement is better. For these computations the effluent BOD at station 5 was set at 13 mg/l or 90% removal of BOD across the secondary grouping. The mixed liquor suspended solids concentration was set at 2000 mg/l. The fraction of suspended solids removed in the primary settler was (.5). All other design parameters used are shown in Table I.

Figure 23 shows the cost of the various groupings of processes as computed by the program for a 10 mgd plant. Assumptions made are the same as those mentioned above. The sludge handling group is the most costly with the secondary group next most costly. The cost of chlorination is included in the plant costs shown in Figures 20 and 22.

Figure 24 shows the total treatment cost for individual processes in the primary group. Cost of chlorination is also shown. Operating and maintenance cost is charged to preliminary treatment and the primary settler but not to the control house. Operating and maintenance cost charged to the chlorination process includes only the cost of chlorine.

PRIMARY TREATMENT PLANTS

Capital Cost, Operating & Maintenance Cost, Debt Services
vs.
Design Capacity

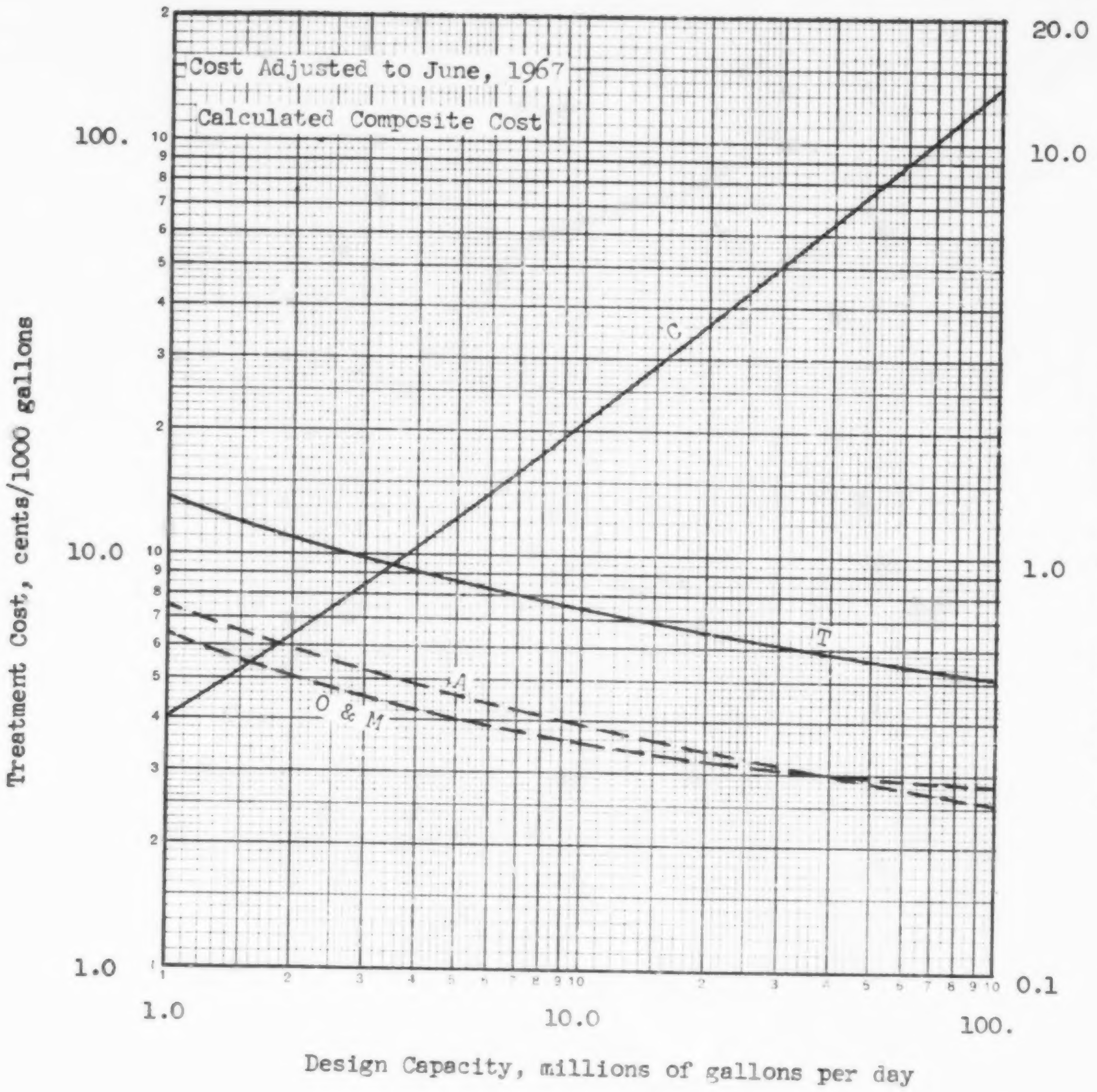


C = Capital Cost, millions of dollars
A = Debt Service, cents per 1000 gallons(4 1/2% - 25 yr.)
O & M = Operating & Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

Figure 19

PRIMARY TREATMENT PLANTS

Capital Cost, Operating & Maintenance Cost, Debt Service
vs.
Design Capacity



C = Capital Cost, millions of dollars
A = Debt Service, cents per 1000 gallons(4 1/2% - 25 yr.)
O & M = Operating & Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

Figure 20

ACTIVATED SLUDGE PLANTS

Capital Cost, Operating & Maintenance Cost, Debt Service
vs.
Design Capacity

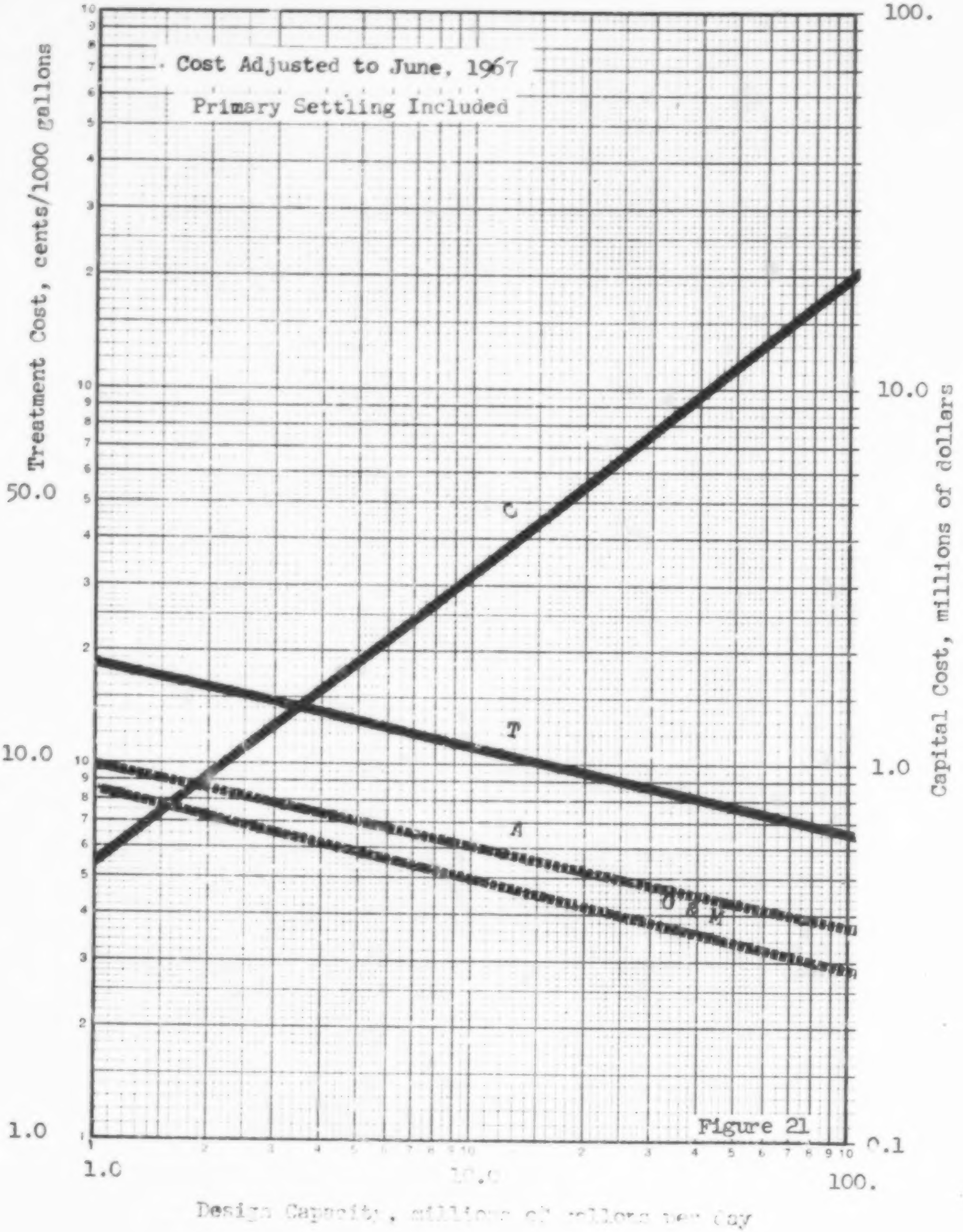
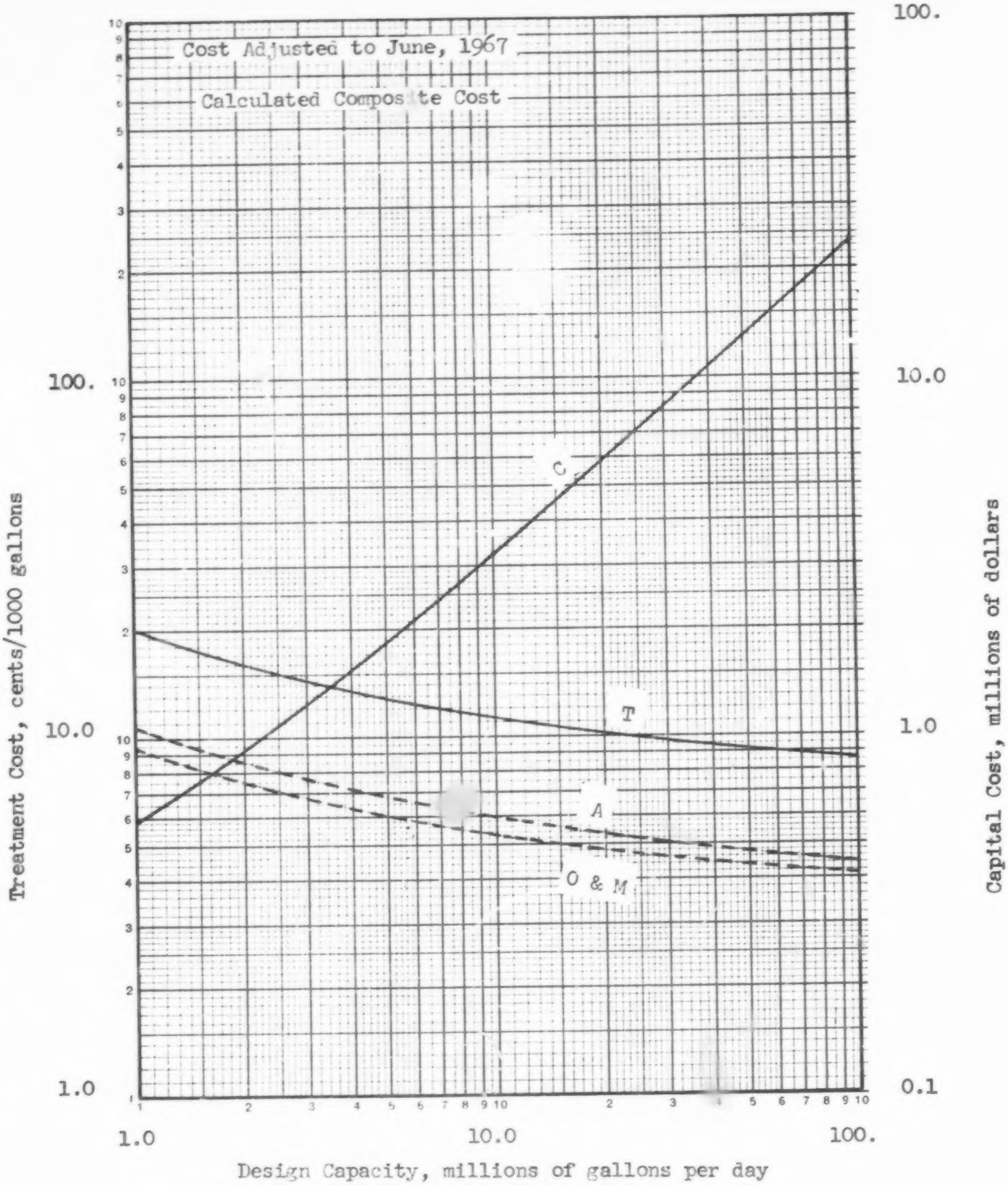


Figure 21

C = Capital Cost, millions of dollars
A = Debt Service, cents per 1000 gallons(4 1/2% - 25 yr.)
O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

ACTIVATED SLUDGE PLANTS

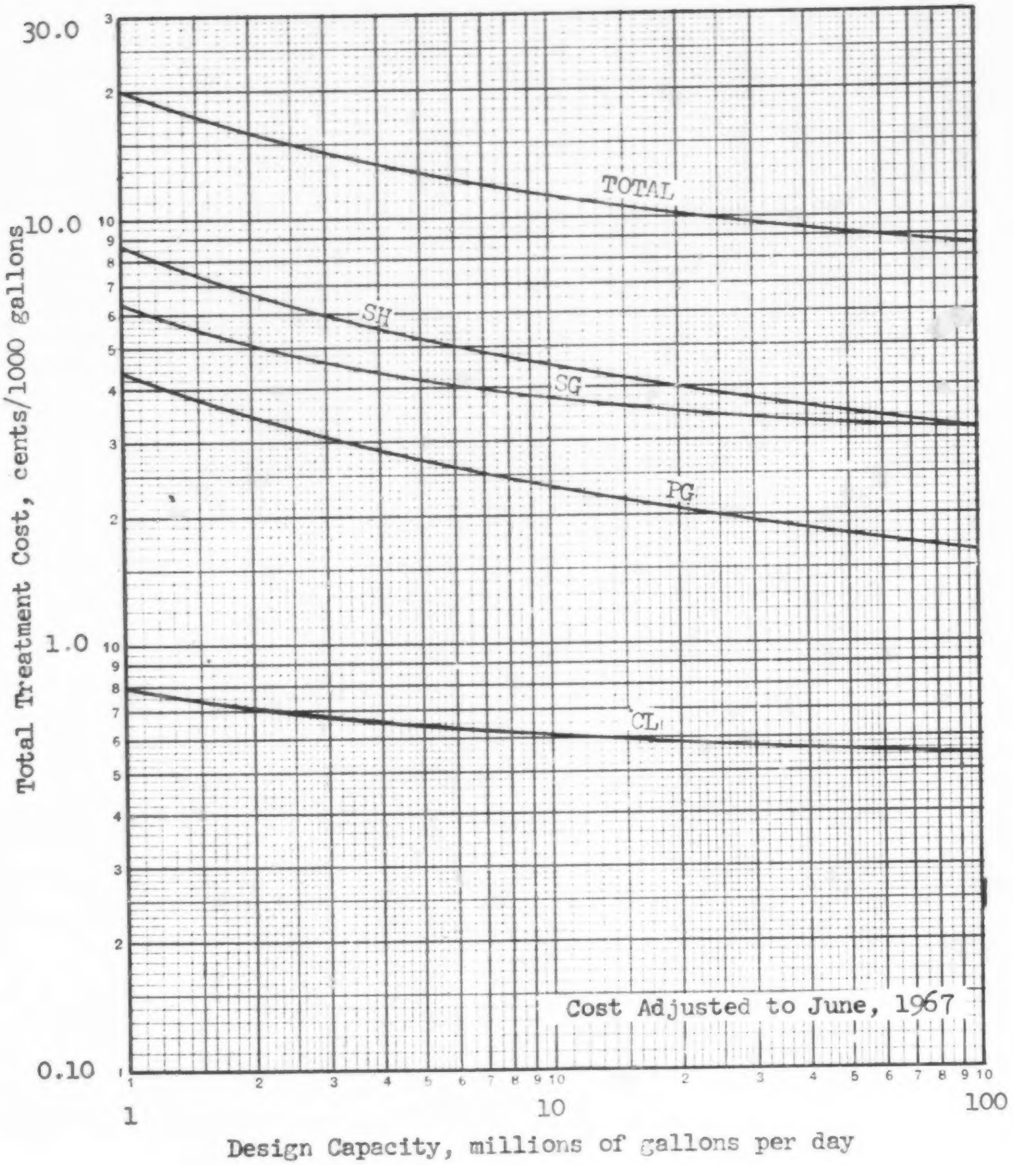
Capital Cost, Operating & Maintenance Cost, Debt Service
vs.
Design Capacity



C = Capital Cost, millions of dollars
A = Debt Service, cents per 100 gallons(4 1/2% - 25 yr.)
O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

Figure 22

CALCULATED COMPOSITE COST-ACTIVATED SLUDGE PLANTS
TOTAL TREATMENT COST ASSOCIATED WITH GROUPS OF PROCESSES
versus
DESIGN CAPACITY

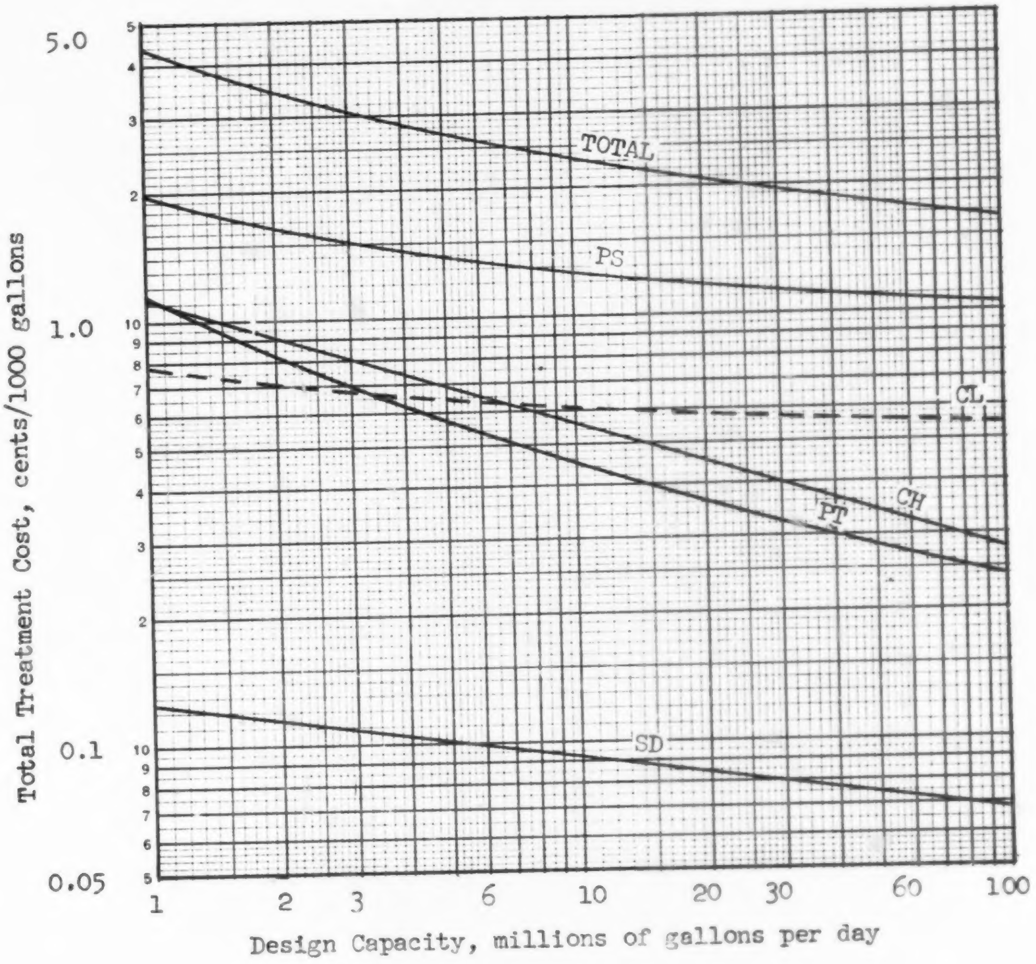


CONDITIONS:
1. 50% Removal of Suspended Solids in Primary Settler
2. Mixed Liquor Suspended Solids in Aerator = 2000 mg/l
3. 90% BOD Removal in Secondary Grouping

LABELS:
Total - Total Plant Cost Including Chlorination
SH - Sludge Handling Grouping
SG - Secondary Treatment Grouping
PG - Primary Treatment Grouping
CL - Chlorination

Figure 23

TOTAL TREATMENT COST FOR INDIVIDUAL PROCESSES
IN THE PRIMARY GROUPING
versus
DESIGN CAPACITY



CONDITION: 50% Removal of Suspended Solids in Primary Settler

LABELS: Total - Total Cost of 4 Items Excluding Chlorination
PS - Primary Settler
CL - Chlorination
SD - Site Development
CH - Control House
PT - Preliminary Treatment

Figure 24

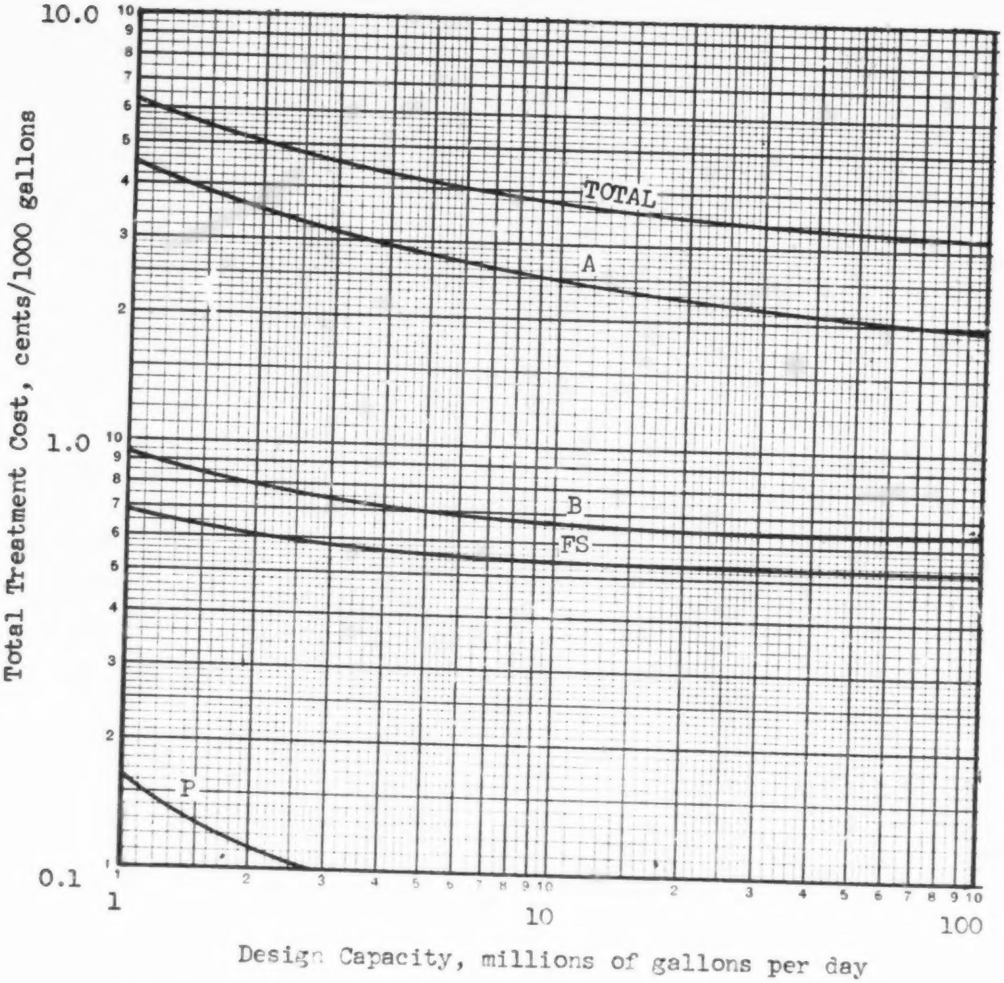
Figure 25 shows the total treatment cost for individual processes in the secondary group. Operating and maintenance cost for the aerator, final settler, and blowers is charged to the aerator. The cost of electrical power is charged to the blowers.

Figure 26 shows the total treatment cost for individual processes in the sludge handling group. No operating and maintenance cost is charged to the thickener and the elutriation process. The cost of chemicals is charged to the vacuum filter. No credit is given for methane produced by the digester.

Figure 2 shows the effect on plant cost of varying the size of the primary or more specifically the fraction of suspended solids removed by the primary settler. The cost of the primary grouping increases without limit as the assumed performance limit of the primary settler is approached. The size of the secondary grouping decreases as the BOD load is reduced. The cost of the sludge handling grouping is affected only slightly. By adding the cost of the primary and secondary groupings a relatively flat optimum region is found between 30 and 50% removal of suspended solids by the primary settler.

Figure 28 shows the theoretical advantage of holding a higher concentration of mixed liquor suspended solids in the aerator. The cost of the secondary grouping is reduced from about 5.8 cents per thousand gallons to about 2.7 cents per thousand gallons by increasing the MLSS from 1000 mg/l to 3800 mg/l, which for a 10 mgd plant corresponds to a savings of about 25%. It should be noted that the possibility of holding higher solids in the aerator is closely related to the settling characteristics of the sludge which is estimated in the program by the empirical relationship expressed as equation (12).

TOTAL TREATMENT COST FOR INDIVIDUAL PROCESSES IN THE SECONDARY GROUPING versus DESIGN CAPACITY

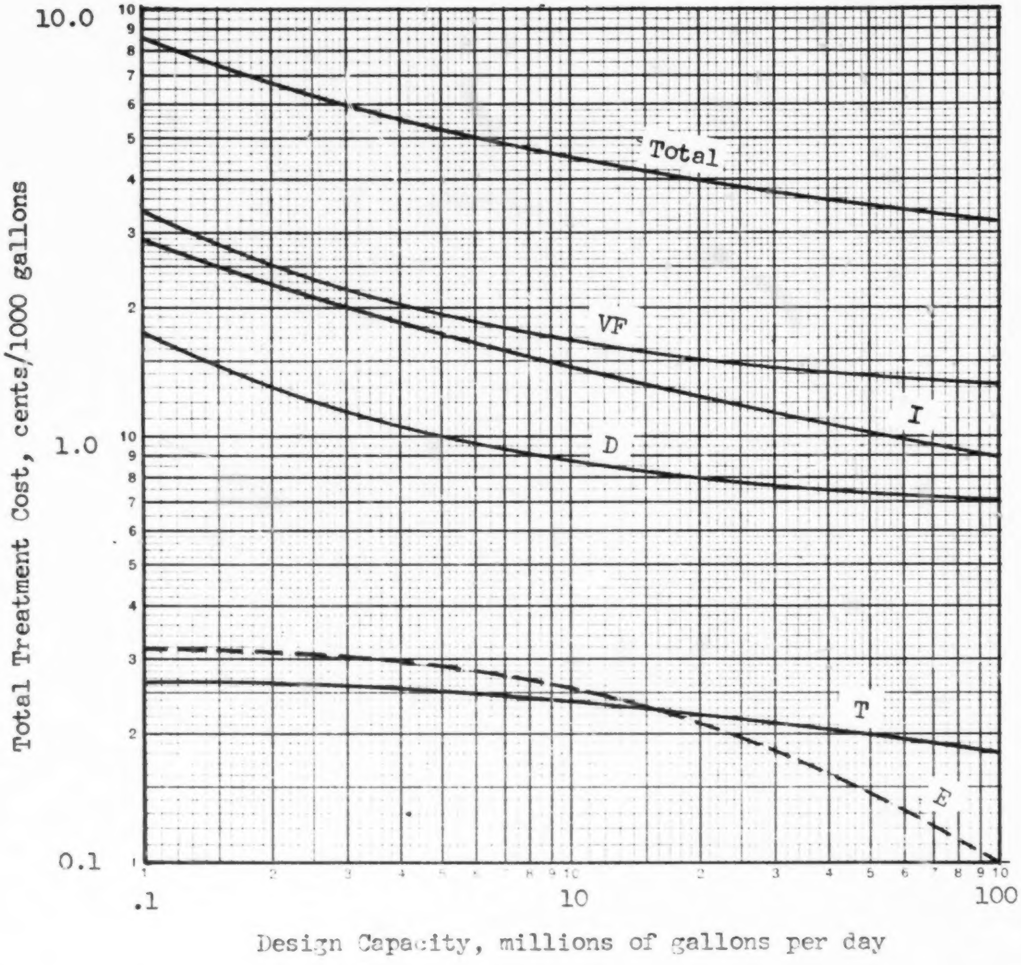


CONDITIONS:
1. BOD Concentration into Aerator = 130 mg/l
2. Mixed Liquor Suspended Solids in Aerator = 2000 mg/l
3. 90% BOD Removal in Secondary Grouping

LABELS: Total - Total Cost of 4 Secondary Components
A - Aerator
B - Air Blowers
FS - Final Settler
P - Sludge Return Pumps

Figure 25

TOTAL TREATMENT COST FOR INDIVIDUAL PROCESSES
IN THE SLUDGE HANDLING GROUP
versus
DESIGN CAPACITY



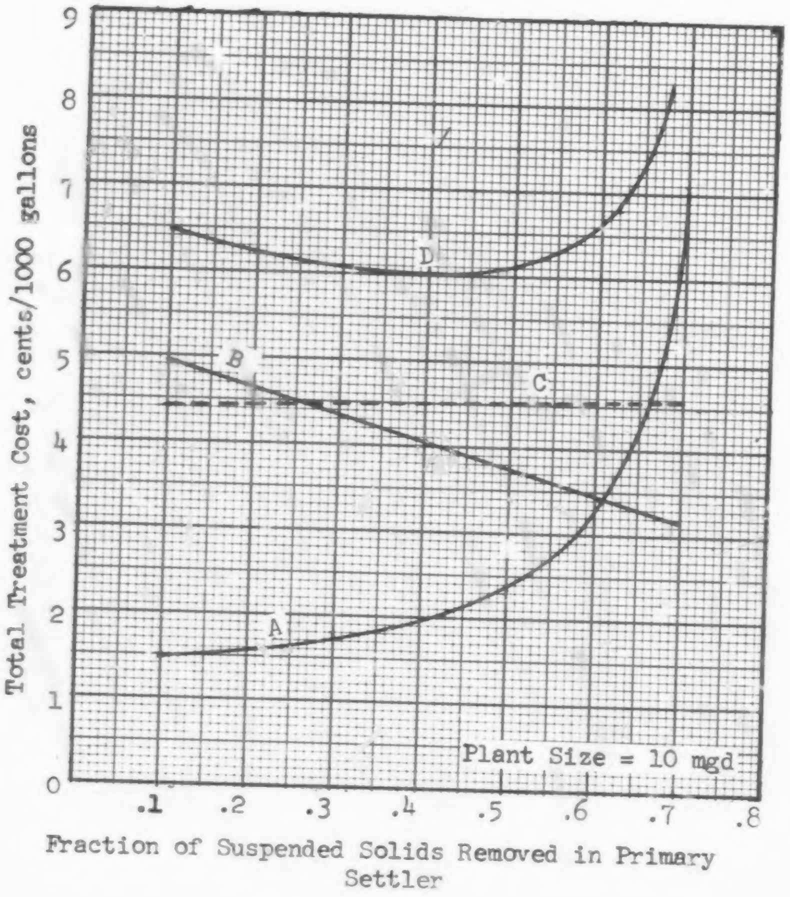
CONDITIONS:
1. 50% Removal of Suspended Solids in Primary Settler
2. Mixed Liquor Suspended Solids in Aerator = 2000 mg/1
3. 90% BOD Removal in Secondary Grouping

LABELS:

Total	Total Cost of All 5 Sludge Handling Processes
I	Incineration
E	Elutriation
VF	Vacuum Filtration
D	Digester
T	Thickener

Figure 26

TOTAL TREATMENT COST FOR VARIOUS PROCESS GROUPINGS
versus
FRACTION OF SUSPENDED SOLIDS REMOVED
IN THE PRIMARY SETTLER



CONDITIONS: MLSS equals 2000 mg/l
BOD at Station No. 5 equals 13 mg/l

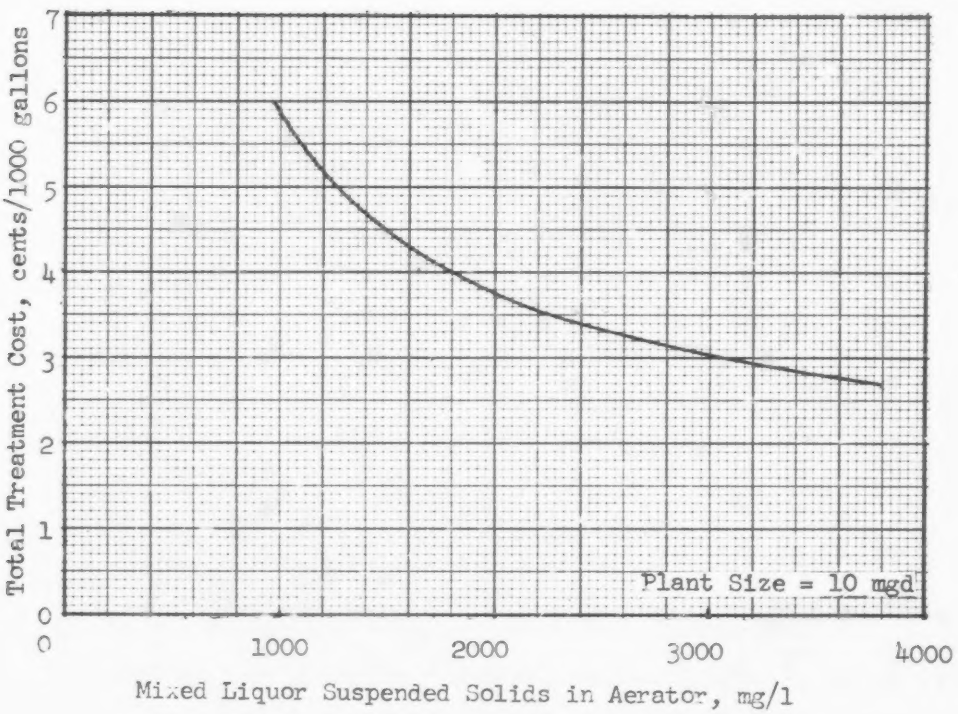
LABELS: A - Primary Grouping
B - Secondary Grouping
C - Sludge Handling Grouping
D - Sum of A and B

Figure 27

TOTAL TREATMENT COST FOR SECONDARY GROUPING

versus

MLSS HELD IN AERATOR



CONDITIONS: 1. BOD achieved at Station 5 equals 13 mg/l
2. FRPS equals 0.5

Figure 28

Figure 29 shows the cost of the secondary groupings as a function of the demand BOD at the effluent of the plant. These differences seem small when compared to the BOD load downstream of the plant. For example, with a MLSS of 2000 mg/l a 90% removal of BOD or an effluent BOD of 13 mg/l costs about 3.8 cents per 1000 gallons while doubling the effluent to 26 mg/l reduces the cost by only 1.5 cents per 1000 gallons to 2.3 cents per 1000 gallons. This represents a savings of only about 13% of total plant cost. The sharp increase in cost represented by the broken line is caused by the occurance of nitrification and the corresponding additional air requirements for the aerator.

Some other interesting relationships have been obtained by use of the computer program. For example, Figures 30, 31, and 32 show the computed distribution of various classes of solids held in the aerator as a function of the operating decisions for the aerator. Generally speaking, the active solids vary with the demand BOD of the effluent and the total concentration of solids held in the aerator. Active solids are generally less than 50% of the total. Debris solids appear to be very small. Non-biodegradable solids represent a surprizingly high percentage of the total. The broken line is caused by the assumption that particulate biodegradable solids remaining are 70% solid and 30% dissolved. The inadequacy of this assumption is obvious and it is hoped that some better way of handling this problem can be found.

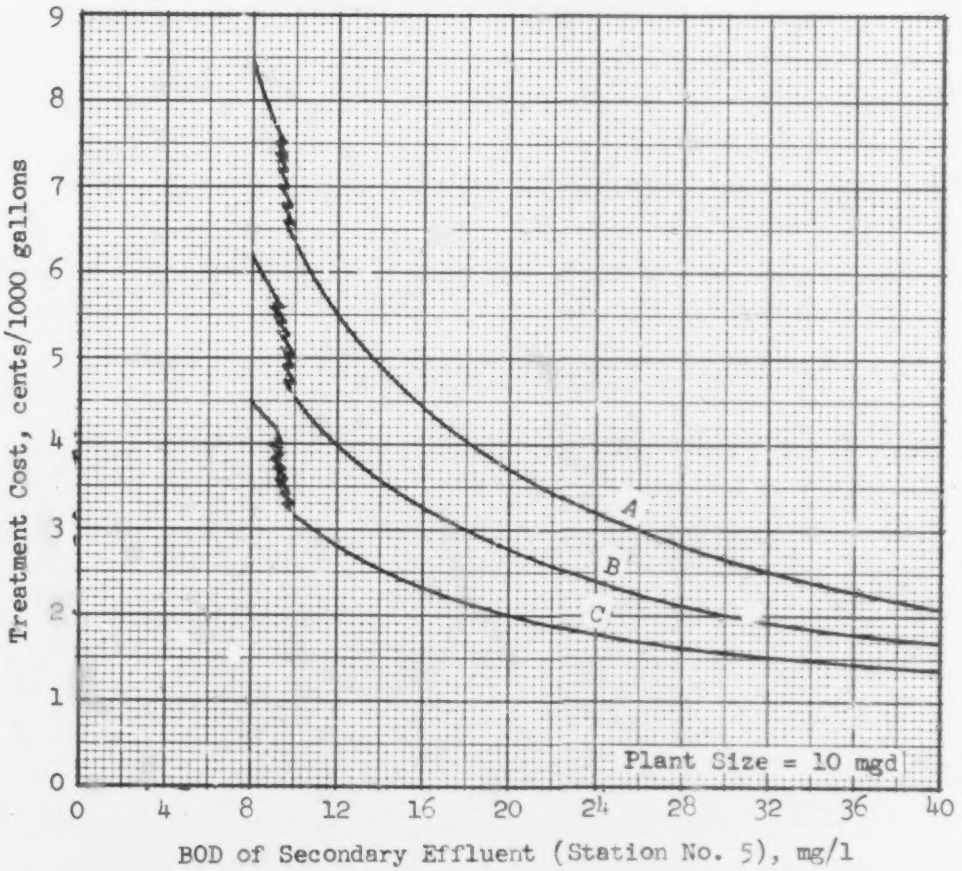
Figure 33 shows the precipitous occurance of nitrification when the BOD demand is increased from 9 to 11 mg/l. The concentration of mixed liquor suspended solids appears to have little effect on the point at which nitrification occurs.

Figure 34 is a map which describes the performance of the secondary group for a 10 mgd plant and influent BOD of 130 mg/l.

TOTAL TREATMENT COST FOR SECONDARY GROUPING

versus

BOD OF SECONDARY EFFLUENT (Station No. 5)

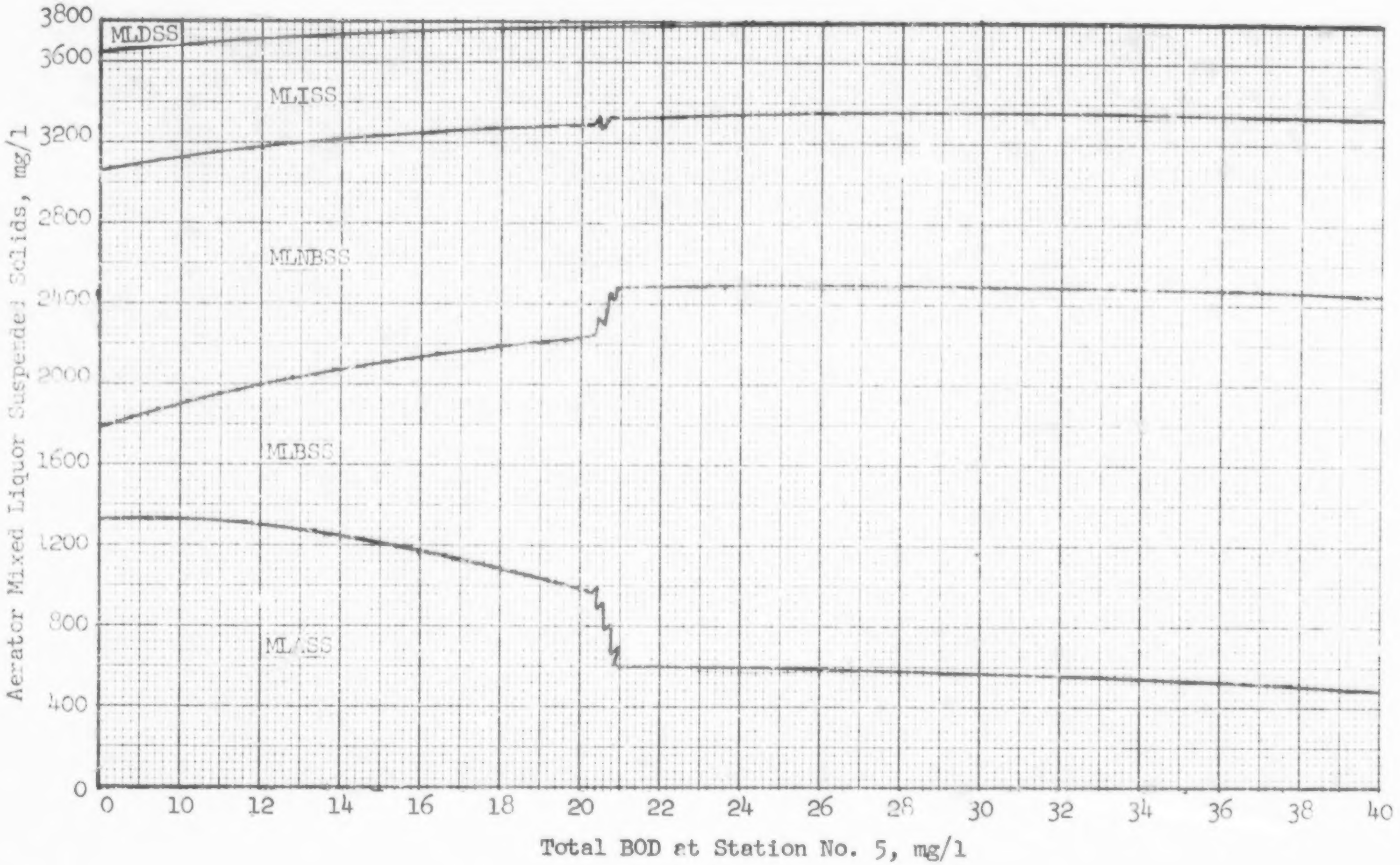


CONDITIONS:
1. FRPS equals 0.5
2. Influent BOD (Station No. 2) equals 130 mg/l

LABELS:
A - Aerator MLSS equals 1200 mg/l
B - Aerator MLSS equals 2000 mg/l
C - Aerator MLSS equals 3800 mg/l

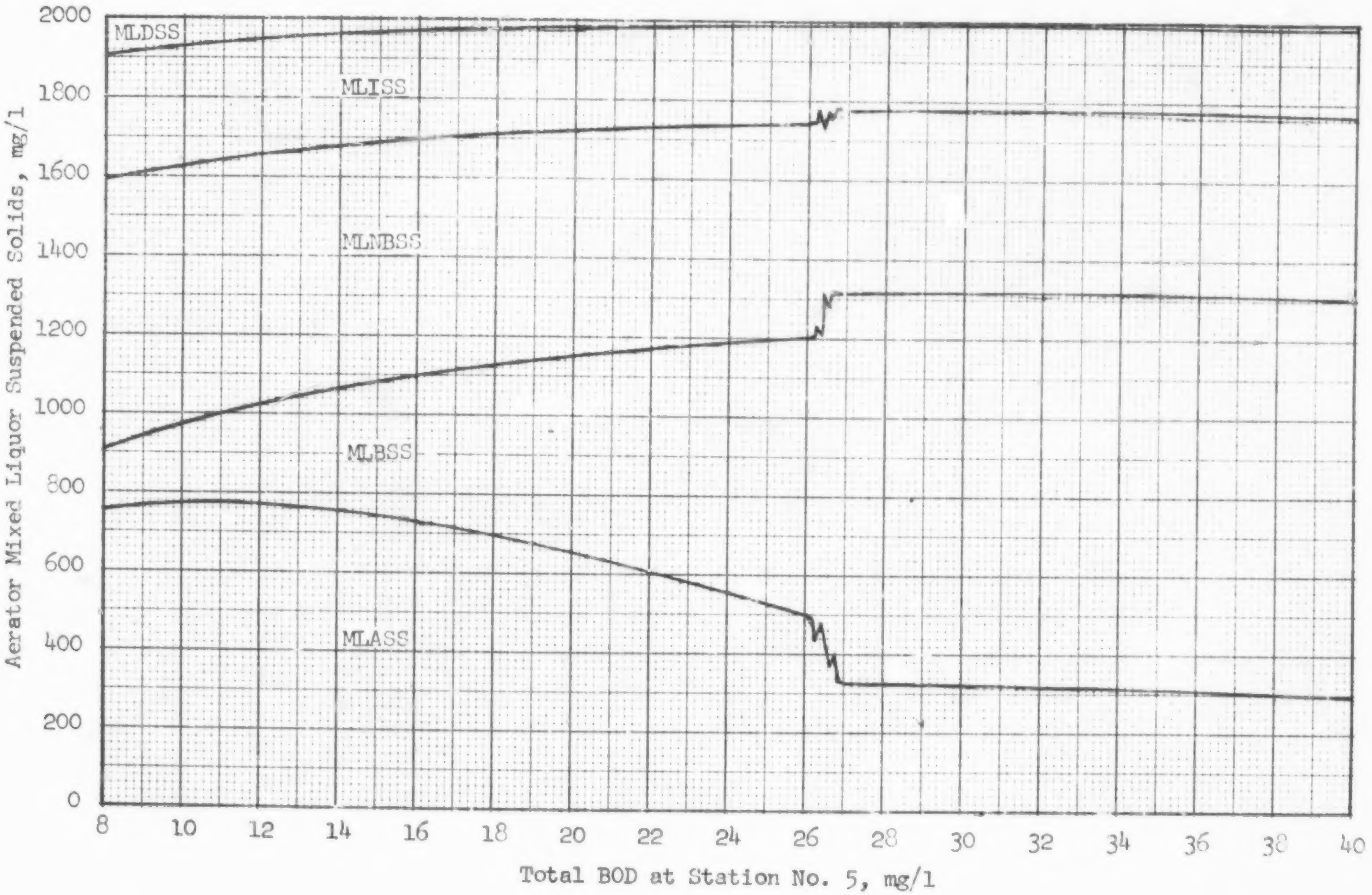
NOTE: Broken Line Indicates That Nitrification Has Occurred

Figure 29



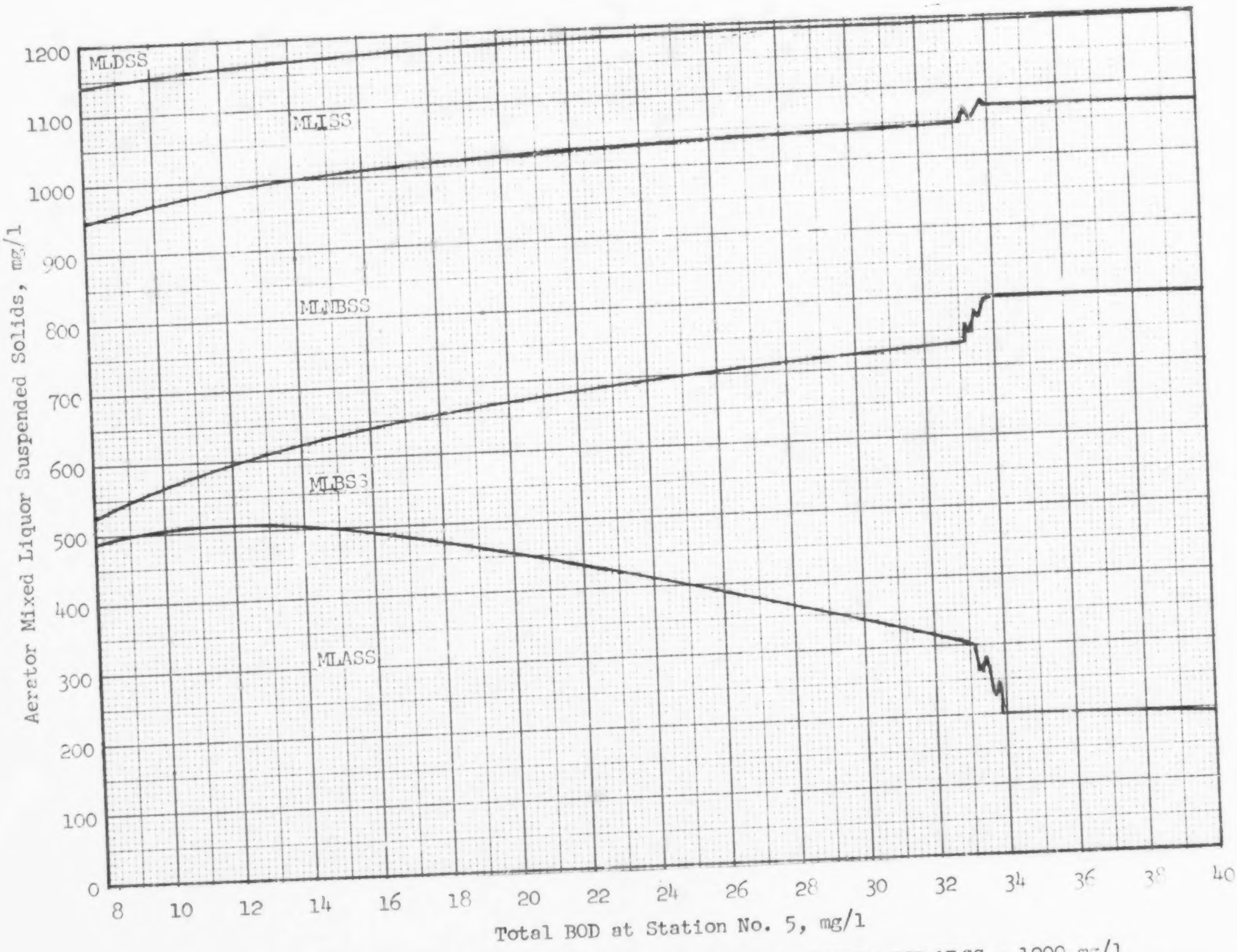
DISTRIBUTION OF AERATOR SOLIDS BETWEEN FIVE ASSUMED CLASSES FOR MLSS = 3800 mg/1

Figure 30



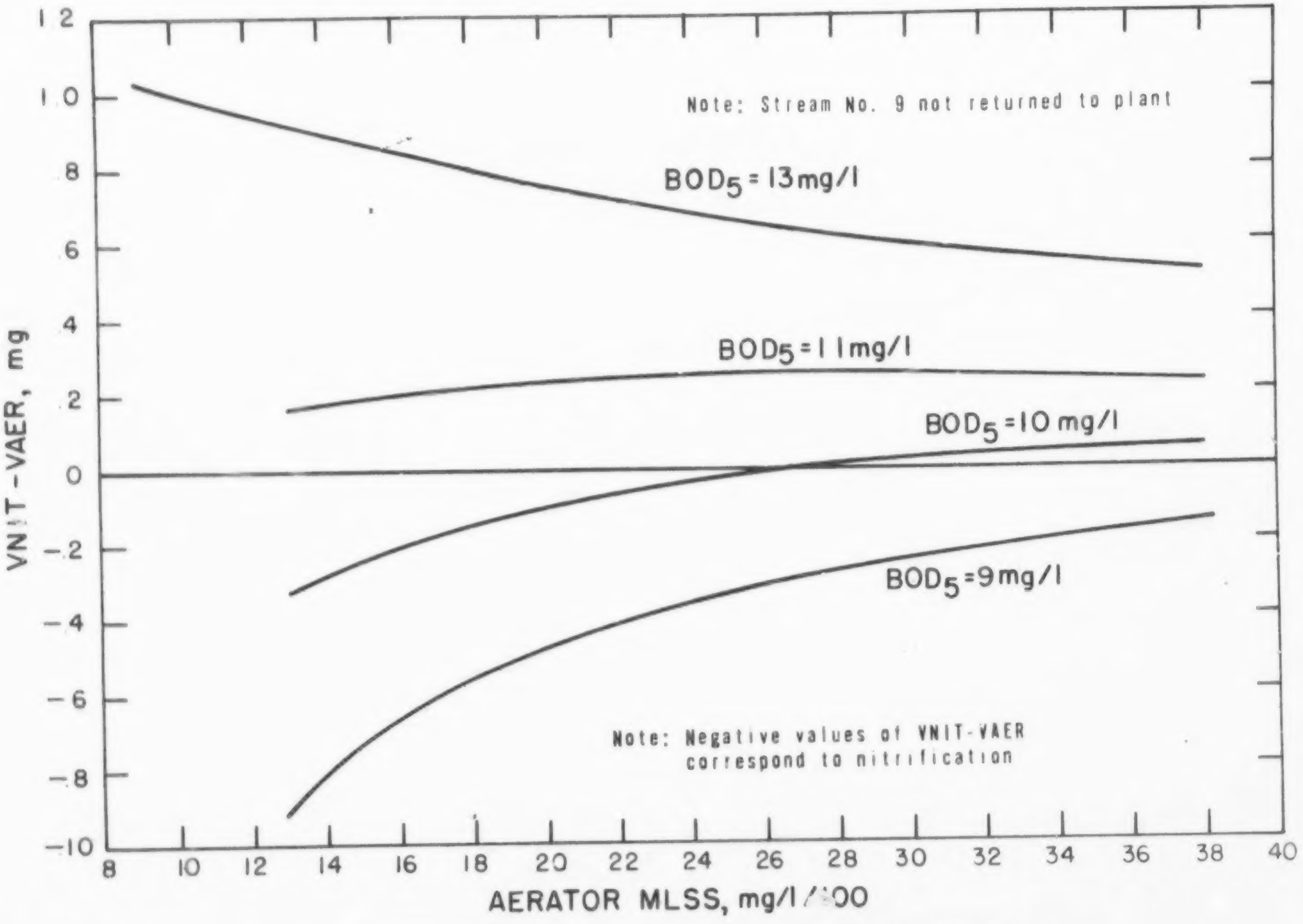
DISTRIBUTION OF AERATOR SOLIDS BETWEEN FIVE ASSUMED CLASSES FOR MLSS = 2000 mg/l

Figure 31



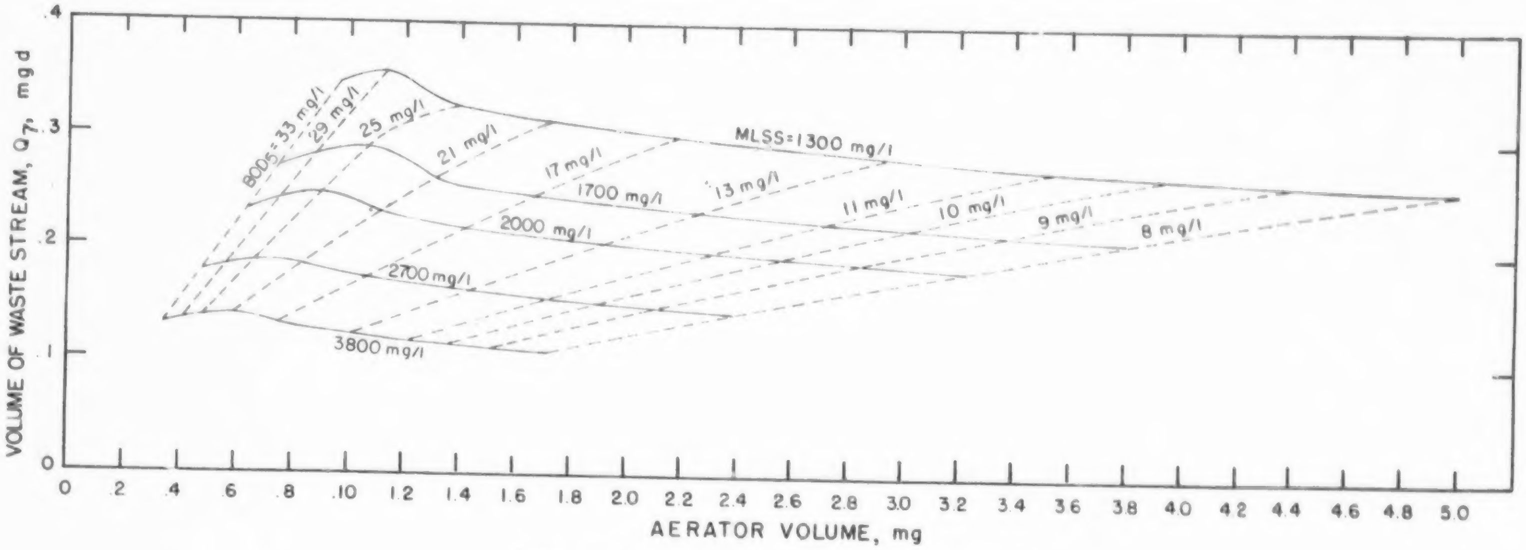
DISTRIBUTION OF AERATOR SOLIDS BETWEEN FIVE ASSUMED CLASSES FOR MLSS = 1200 mg/l

Figure 32



EFFECT OF BOD_5 AND MLSS ON CONDITIONS FOR NITRIFICATION

Figure 33



SLUDGE WASTE STREAM, Q_7 versus MLSS, BOD_5 AND AERATOR VOLUME

Figure 34

COMPARISON WITH MEASUREMENT ON FULL-SIZED PLANT

Assumptions made in developing the program have been compared to measurements made at only one plant. In connection with a study of the effectiveness of adding polyelectrolyte to the primary settler of the Washington, D.C. Water Pollution Control Plant a series of baseline 24 hour composite samples were taken at the plant before any polyelectrolyte was added. These measurements were made each day between April 3, 1967 and April 30, 1967. The measured values shown in Table III are averages over this period. The most important comparisons are the rate constant for the aerator, CAER20, and the fraction of solids in the aerator which escapes over the weirs, XRSS. The average measured value for the rate constant, CAER20, was (0.034) which is 40% higher than the value of (0.024) used in the program. If equation (8) had been used to compute the expected BOD at station 5 for the Washington D.C. plant a value of 54.5 mg/l would have been found which is 23% higher than the actual observed value of 44.3 mg/l.

The value of XRSS computed from equation (12) was (0.165) as compared to an observed value of (0.078). The predicted suspended solids concentration in the effluent would have been 73.3 mg/l as compared to the actual observed value of 34.7 mg/l. This is a substantial discrepancy and emphasizes the importance of learning more about the performance of the final settler.

Table III

COMPARISON OF PARAMETERS ASSUMED OR COMPUTED IN THE PROGRAM

AND

VALUES MEASURED AT THE DISTRICT OF COLUMBIA WATER POLLUTION CONTROL PLANT

Measured Values are Averages of 24 Hour Composite Samples
For the Month of April, 1967

	Measured or Actual Value for Plant	Assumed or Calculated Value from Program
1. Primary Settler Overflow Rate gpd/sq. ft.	16[illegible]8.	
2. Fraction of Suspended Solids Removed in Primary Settler, FRPS	0.66	0.45
3. Concentration Ratio of Solids in Primary Settler, URPS	66.	400.
4. Rate Constant for Aerator, $CAER_{20}$	0.034	0.024
5. Aerator Detention Time, TA, hrs.	2.59	
6. Mixed Liquor Suspended Solids, mg/l	444.	
7. Fraction of Solids Over Final Settler Weirs, XRSS	0.078	0.165
8. Concentration Ratio for Solids in Final Settler, URSS	46.5	3.
9. Final Settler Overflow Rate gpd/sq. ft.	969.	
10. Solids Recovery Ratio for Sludge Thickener	1.24*	.95
11. Digester Detention Time, days	21.56	
12. Fraction of Volatile Solids Destroyed in Digester	.706	.60 in 15 days
13. Solids Recovery Ratio for Elutriation, ERR	.717	.76
14. Wash Water Ratio for Elutriation, WRE	4.24	3.

* Due to measurement error

LIMITATIONS, OMISSIONS, AND RECOMMENDATIONS FOR ADDITIONAL WORK

The model which has been presented for a conventional water renovation system is viewed as a first attempt to represent the performance and cost as a function of the characteristics of the feed stream and operating and design decisions associated with each process. The program has been slanted towards preliminary design rather than towards simulation but, since only steady state performance is considered, this difference is not fundamental.

A key point which cannot be over emphasized is that this work represents a beginning and not a finished product which can be used with confidence by planners or by design engineers. The second phase of the work will be to make measurements at a full-sized activated sludge plant which is capable of operating over wide ranges of design parameters, and to compare the performance of the plant with the mathematical model. This will, undoubtedly, result in adjustment of parameters or perhaps even major changes in the models for individual processes. Some of the questions which will, hopefully, be answered by such a measurement program are given below, together with the corresponding assumption made in the mathematical model.

1. Can the biodegradable fraction of carbon entering the plant be evaluated by multiplying the 5-day BOD by an appropriate factor:
   ASSUMPTION: The factor is assumed to be 1.87 in the program.

2. Does the primary settler have the same removal efficiency towards all classes of suspended solids, or is a more complicated representation necessary?
   ASSUMPTION: Removal efficiency assumed to be the same for all classes of solids.

3. How is the removal efficiency of each class of solids related to the overflow rate of the primary settler?
ASSUMPTION: An exponential relationship is used, based on the observed removal of suspended solids from many plants.

4. How is the density of the underflow sludge in the primary settler related to the pumping rate and the storage volume available in the settler?
ASSUMPTION: Concentration of each class of solids is assumed equal to a constant times the influent concentration.

5. How can the rate of synthesis of dissolved versus particulate BOD be compared?
ASSUMPTION: Dissolved BOD is used before any particulate BOD is used.

6. What is the rate constant and stoichiometry involved in the synthesis of active solids from biodegradable carbon?
ASSUMPTION: Rate constant is deduced from average measurements made across aerator and final settler from full-sized plants. 0.65 lb active solids produced per lb of BOD synthesized.

7. Measure the rate of lysis of active solids in the aerator and determine the temperature dependency.
ASSUMPTION: 18% of active solids lyse each day.

8. Must particulate BOD be hydrolyzed before it can be used as food, and what are the relative rates of hydrolysis and synthesis?
ASSUMPTION: It is assumed that unsynthesized particulate food is 90% solid and 10% dissolved.

9. How does the amount of suspended solids which escapes over the final settler weirs depend on process parameters such as MLSS, aerator detention time, or overflow rate of the final settler?
ASSUMPTION: Empirical relationship from bench scale tests used.

10. How does the density of sludge pumped from the final settler depend on the sludge pumping rate and the storage capacity of the settler?
ASSUMPTION: Density taken as independent of pumping rate and equal to a constant times MLSS.

11. How does the removal efficiency of the final settler depend on the nature of the solids?
ASSUMPTION: Removal efficiency assumed the same for all classes of solids.

12. Nitrification must be considered complete when some arbitrary fraction of the ammonia has been converted to nitrite. What fraction should be used?
ASSUMPTION: 99% is used in the program.

13. How does the relationship assumed for oxygen usage in the aerator agree with observed results?
ASSUMPTION: In the program it is related to the food used and the active solids produced each day.

14. Does the rate of synthesis of active solids depend on the dissolved oxygen level in the aerator?
ASSUMPTION: In the program the rate of synthesis is taken as independent of the DO level but some evidence exists which suggests a dependency.

15. Does the additional oxygen demand caused by nitrification agree with computed results?
ASSUMPTION: Downing's work used in the program.

16. How does the degree of thickening accomplished in the thickener depend on the overflow rate or the solids loading rate?
ASSUMPTION: No such relationship used. Average design point data used.

17. How can the solids concentration in the overflow stream from the thickener be related to the characteristics of the stream into the thickener?
ASSUMPTION: TSS_{11} is computed from assumed values for TRR and TSS_{12}.

18. Is it correct to take only biodegradable carbon entering the digester as being capable of decomposition in the digester?
ASSUMPTION: This is assumed in the program.

19. Evaluate the rate of decomposition in the digester as a function of the detention time and the sludge temperature.
ASSUMPTION: Results of McCarty used in program.

20. How can the increase in alkalinity across the digester be related to the amount of decomposition accomplished in the digester?
ASSUMPTION: All but 4.2% of the released nitrogen assumed converted to ammonia with corresponding increase in alkalinity.

21. How much nitrogen escaped from the digester as nitrogen gas?
ASSUMPTION: 4.2% assumed.

22. How can the solids recovery ratio and the underflow solids concentration of the elutriation process be related to the size of the structure?
ASSUMPTION: No relationship used in the program. Design point data used.

23. How is the performance of the vacuum filter related to the amount of chemicals used and the properties of the sludge? ASSUMPTION: Average values used. No relationship available.

After completion of the second phase of intensive comparison between model and full-sized plant performance, several additional plants should be compared to the model in a less exhaustive way until the required degree of confidence in the model is established.

Several additions should be made to the program which will make it more useful to plant designers, for example, the heat requirements for the digester or the use of sludge gas as fuel for internal combustion engines or as heating fuel. The model for the sludge incineration process is also very incomplete. Capital and operating cost data are lumped in the program as it now stands and no attempt has been made to compute the products of combustion, both gaseous and solid. Several alternative methods of sludge oxidation such as the atomized suspension technique, evaporation followed by incineration, or wet oxidation should be included. Ultimate disposal methods such as barging to sea, pipelines, truck or rail transportation to landfills should be added to the program.

Another severe limitation of the program shown in the Appendix is the fixed configuration of the processes. This has been remedied by development at the Cincinnati Water Research Laboratory of a calling program which allows the performance and cost of any group of the eight processes shown in Figure 1 to be computed when arranged in any configuration. This program was patterned after the PACER program which was developed by Mosler[71] working under Dr. Paul T. Shannon at Purdue University.

Finally, a need exists for more valid and up-to-date relationships for the capital cost of structures and equipment, and particularly for operating cost which includes the cost of labor, chemicals, and power.

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68. Rowan, P. P., Jenkins, K. L., and Howells, D. H., "Estimating Sewage Treatment Plant Operation and Maintenance Costs," Journal Water Pollution Control Federation, Vol. 33, 1961, pp. 111-121.

69. Swanson, C. L., "Unit Process Operating and Maintenance Costs for Conventional Sewage Treatment Processes," Internal FWPCA Memorandum, August 1, 1966.

70. Smith, Robert, "A Compilation of Cost Information for Conventional and Advanced Wastewater Treatment Plants and Processes," Internal FWPCA Report December, 1967.

LIMITATIONS, OMISSIONS, AND RECOMMENDATIONS FOR ADDITIONAL WORK

71. Mosler, H.A., "PACER - A Digital Computer Executive Program for Process Simulation and Design," M.S. Thesis, Purdue University, Sept., 1963.

APPENDIX

Table A-1 VARIABLES AND PARAMETERS USED IN COST-PERFORMANCE COMPUTATION

Text Symbols	Computer Symbols	Nominal Value	Definitions and Comments
	ACOST(j)		Amortization cost for the jth process, dollars per year.
	AE		Surface area of elutriation tank, sq. ft.
AEFF	AEFF		Efficiency of diffusers in aerator corrected for water temperature and dissolved oxygen deficit.
$AEFF_{20}$	AEFF20	0.075	Efficiency of diffusers in aerator at zero dissolved oxygen and 20°C.
	AF	0.0736	Vacuum filter area, sq. ft.
	AFS		Surface area of final settler, sq. ft./1000.
AIRCFD	AIRCFD		Air requirements for aerator in standard cubic feet per day.
ALK_1	ALK(i)		Concentration of alkalinity as $CaCO_3$ at Station No. 1, mg/l[a].
	AOCOST(j)		Sum of operating and amortization cost for the jth process, dollars per year.
APS	APS		Surface area of primary settler, sq. ft./1000.
	ASB		Area of sludge drying beds, sq. ft.
	ASMAX		Current maximum value of MLASS, mg/l mass.
	ASMIN		Current minimum value for MLASS, mg/l mass.
	ASS(i)		The ith input value for MLSS.
	ASS(k)		Input values for mixed liquor suspended solids to be held in the aerator.
	ATH1		Surface area of thickener based on overflow rate, sq. ft.
	ATH2		Surface area of thickener based on solids loading rate, sq. ft.

[a] mg/l = milligrams per liter.

Table A-1 VARIABLES AND PARAMETERS USED IN COST-PERFORMANCE COMPUTATION (Cont.)

Text Symbols	Computer Symbols	Nominal Value	Definitions and Comments
	ATHM		Greater value of ATH1 and ATH2, sq. ft.
	AVF		Area of vacuum filter, sq. ft.
	BODI		Total BOD[a] at Station No. I, mg/l oxygen.
	BODREM		Fractional 5-day BOD removal across aerator.
	BSIZE		Required size of blower for supplying air to aerator, cubic feet per minute.
$CAER_{20}$	CAER20	1.0	Rate constant used for sizing the aerator when water temperature is 20°C.
CAER	CAER		Rate constant for sizing aerator corrected for water temperature.
	CAIRP		Cost of electrical power for operating blowers, dollars per year.
	CCI	1.25	Capital cost index to account for the variation of construction cost with time.
	CCOST(j)		Capital cost of the jth process in dollars.
	CEDR	0.18	Rate at which active solids are destroyed by natural causes in the aerator, fraction/day mass.
	CFECL3	0.08	Cost of ferric chloride for treating washed sludge, dollars per pound.
	CFPGAL		Air requirement for aerator, scf[b]/gallon of sewage at Station No. 1.
	CH4CFD		Standard cubic feet of methane produced in the digester each day, scf/day methane.
	CKWH	0.01	Cost of electrical power, dollars/kilowatt-hour.

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[a] BOD = Biochemical Oxygen Demand.

[b] scf = Standard cubic feet.

Table A-1 VARIABLES AND PARAMETERS USED IN COST-PERFORMANCE COMPUTATION (Cont.)

Text Symbols	Computer Symbols	Nominal Value	Definitions and Comments
	CNIT		Rate constant used in nitrification calculation.
	CO2CFD		Standard cubic feet of carbon dioxide produced in the digester each day, scf/day CO_2.
	COSTO(i)		Operating cost for the ith process, dollars/year.
	COSTO(j)		Operating cost for the jth process, dollars/year.
	CPERKG(i)		Total cost of the ith process, cents/1000 gal.
	CREM		Fraction of organic carbon removed across the aerator.
K_1	C1DIG	0.25	Rate constant for digester.
K_2	C2DIG	447.5	Rate constant for digester.
	CWR		Carbon waste ratio, ratio of carbon in waste stream (Station 7) to carbon in influent stream (Station 2), (lb/day)/(lb/day).
$DBOD_i$	DBOD(i)		Dissolved BOD at Station No. i.
	DBOD(4)		Dissolved unmetabolized 5-day BOD in aerator, mg/l[a] oxygen.
T	DEGC	20.0	Water temperature, degrees Centigrade.
	DEMBOD(i)		The ith input value for BOD_5.
	DEMBOD(k)		Input values for demand value of 5-day BOD at Station No. 5.
DFM_i	DFM(i)		Dissolved inorganic species at Station No. i, mg/l mass.
$DIGC_{12}$	DIGC12		Concentration of biodegradable carbon at Station 12, mg/l carbon.

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[a] mg/l = milligrams per liter.

Table A-1 VARIABLES AND PARAMETERS USED IN COST-PERFORMANCE COMPUTATION (Cont.)

Text Symbols	Computer Symbols	Nominal Value	Definitions and Comments
$DIGC_{13}$	DIGC13		Concentration of biodegradable carbon at Station 13, mg/l[a] carbon.
DN_i	DN(i)		Concentration of dissolved nitrogen at Station No. i, mg/l nitrogen.
$DNBC_i$	DNBC(i)		Concentration of dissolved non-biodegradable carbon at Station No. i, mg/l carbon.
DO	DO	1.0	Concentration of dissolved oxygen in aerator, mg/l oxygen.
DOC_i	DOC(i)		Concentration of all dissolved organic carbon at Station No. i, mg/l carbon.
DOSAT	DOSAT		Saturation value for dissolved oxygen in the aerator at one-half the water depth, mg/l oxygen.
DP_i	DP(i)		Concentration of dissolved phosphorus at Station No. i, mg/l phosphorus.
	EFF		Efficiency of aerator, pounds of carbon which escape to the atmosphere per day as carbon dioxide divided by pounds of carbon entering aerator per day.
ERR	ERR	0.76	Solids recovery ratio for elutriation, total solids in stream 15 divided by total solids in stream 13, (lb/day)/(lb/day).
	ERROR		Difference between FMAX and FMIN.
FECL3	$FECL_3$		Concentration of ferric chloride used for sludge conditioning, mg/l $FeCl_3$.
$FOOD_{max}$	FMAX		Current maximum value for FOOD, mg/l 5-day BOD[b].
$FOOD_{min}$	FMIN		Current minimum value for FOOD, mg/l 5-day BOD.

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[a] mg/l = milligrams per liter.

[b] BOD = Biochemical Oxygen Demand.

Table A-1 VARIABLES AND PARAMETERS USED IN COST-PERFORMANCE COMPUTATION (Cont.)

Text Symbols	Computer Symbols	Nominal Value	Definitions and Comments
FOOD	FOOD		5-day BOD[a] synthesized to active solids in the aerator each day, mg/l[b] oxygen.
FRDIG	FRDIG		Fraction of carbon entering the digester which is converted to gas and removed from the sludge stream.
FRPS	FRPS	0.5	Fraction of solids entering the primary settler which is removed from the main stream and sent to the thickener via stream 8.
GE	GE	800.0	Design overflow rate for elutriation, gpd[c]/sq. ft.
GES	GES	9.0	Design solids loading rate for elutriation, lb/day/sq. ft.
GPS	GPS		Computed overflow rate for primary settler, gpd/sq. ft.
GSS	GSS	2000.0	Design overflow rate for final settler, gpd/sq. ft.
GSTH	GSTH	9.0	Design solids loading rate for thickener, lb/day/sq. ft.
GTH	GTH	750.0	Design overflow rate for thickener, gpd/sq. ft.
	LOOPS		Number of iterative recycle calculations when Stream No. 9 is returned.
	NAS		Number of input values for MLSS.
	NBOD		Number of input values for BOD_5.
	NFORK(i)		An input parameter used in the program to include or exclude various processes.

---

[a] BOD = Biochemical Oxygen Demand.

[b] mg/l = Milligrams per liter.

[c] gpd = Gallons per day.

Table A-1 VARIABLES AND PARAMETERS USED IN COST-PERFORMANCE COMPUTATION (Cont.)

Text Symbols	Computer Symbols	Nominal Value	Definitions and Comments
	NFORK(j)		An input parameter which takes a value of 0 or 1 and is used in the program to include or exclude various processes.
Q_i	Q(i)		Volume flow at Station No. i, mgd[a].
	REMN		Fraction of nitrogen removed from the main stream by the aerator.
	REMP		Fraction of phosphorus removed from the main stream by the aerator.
p	RETURN		Sludge return ratio, Q_6/Q_2.
$SBOD_i$	SBOD(i)		Solid 5-day BOD[b], mg/l[c] oxygen at Station No. i.
	SBOD(4)		Solid unmetabolized 5-day BOD in aerator, mg/l oxygen.
SFM_i	SFM(i)		Concentration of solid inorganic matter at Station No. i, mg/l mass.
$SNBC_i$	SNBC(i)		Concentration of solid non-biodegradable carbon at Station No. i, mg/l carbon.
SOC_i	SOC(i)		Concentration of all solid carbon at Station No. i, mg/l carbon.
SON_i	SON(i)		Concentration of solid nitrogen at Station No. i, mg/l nitrogen.
SOP_i	SOP(i)		Concentration of solid phosphorus at Station No. i, mg/l phosphorus.
	TA		Aerator detention time, VAER/Q_2, days.
t	TAN		Aerator detention time required to achieve nitrification, days.

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[a] mgd = Million gallons per day.

[b] BOD = Biochemical Oxygen Demand.

[c] mg/l = Milligrams per liter.

Table A-1 VARIABLES AND PARAMETERS USED IN COST-PERFORMANCE COMPUTATION (Cont.)

Text Symbols	Computer Symbols	Nominal Value	Definitions and Comments
	TBOD2		Total concentration of 5-day BOD[a] at Station 2, mg/l[b] oxygen.
	TBOD5		Total concentration of 5-day BOD at Station 5, mg/l oxygen.
TD	TD	15.0	Digester detention time, days.
T	TDIG	33.0	Sludge temperature in digester, degrees Centigrade.
	TEMPI		The Ith temporary storage space.
	TOC2		Concentration of all carbon at Station 2, mg/l carbon.
	TOC5		Concentration of all carbon at Station 5, mg/l carbon.
	TOC7		Concentration of all carbon at Station 7, mg/l carbon.
	TOL		Tolerance used in iterative loop.
	TOTAOC		Total amortization and operating cost for all processes, dollars/year.
	TOTCC		Total capital cost for all processes, dollars.
TRR	TRR	0.95	Solids recovery ratio for thickening, total solids in stream 12 divided by total solids in stream 10, (lb/day)/(lb/day).
TSS_1	TSS(i)		Total suspended solids at Station No. i, mg/l mass.
TVF	TVF	0.238	Fraction of the time vacuum filters are operated.
URPS	URPS	400.0	Ratio of solids concentration in stream 8 to solids concentration in stream 1, TSS_8/TSS_1.

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[a] BOD = Biochemical Oxygen Demand.

[b] mg/l = Milligrams per liter.

Table A-1 VARIABLES AND PARAMETERS USED IN COST-PERFORMANCE COMPUTATION (Cont.)

Text Symbols	Computer Symbols	Nominal Value	Definitions and Comments
URSS	URSS	3.0	Ratio of solids concentration in underflow stream from final settler to the total solids concentration in the aerator, TSS_7/MLSS.
VAER	VAER		Volume of aerator, millions of gallons.
VDIG	VDIG		Volume of digester, cubic feet/1000.
VFL	VFL	4.9	Vacuum filter loading, gal/hr/sq. ft.
VNIT	VNIT		Volume of aerator required to achieve nitrification, millions of gallons.
VSS_i	VSS(i)		Concentration of volatile suspended solids at Station No. i, mg/l[a] mass.
	WAS		Pounds of active solids in the aerator, lb mass.
	WDN		Pounds per day of dissolved nitrogen in streams 5 and 7 from the aerator, lb/day nitrogen.
	WFOOD		Pounds per day of 5-day BOD[b] synthesized to active solids in the aerator, lb/day oxygen.
	WP		Percentage of moisture in filtered sludge.
	WRE	3.0	Wash water ratio for elutriation, Q_{17}/Q_{13}.
MLNBSS	XLNBSS		Concentration of inert organic solids in aerator caused by inert organic solids in influent, mg/l mass.
MLASS	XMLASS		Concentration of active solids held in aerator, mg/l mass.
MLBSS	XMLBSS		Concentration of unmetabolized biodegradable solids held in the aerator, mg/l mass.

[a] mg/l = Milligrams per liter.

[b] BOD = Biochemical Oxygen Demand.

Table A-1 VARIABLES AND PARAMETERS USED IN COST-PERFORMANCE COMPUTATION (Cont.)

Text Symbols	Computer Symbols	Nominal Value	Definitions and Comments
MLDSS	XMLDSS		Concentration of non-biodegradable solids in the aerator caused by destruction of active solids by natural causes, mg/1[a] mass.
MLISS	XMLISS		Concentration of inert inorganic solids in aerator, caused by inorganic solids in influent, mg/1 mass.
MLSS	XMLSS		Total concentration of solids in aerator, mg/1 mass.
	XRSS		Ratio of solids in the overflow stream from the aerator to the total solids held, TSS_5/MLSS.
	Y1		Alkalinity of sludge expressed as a percentage of the total solids.
	Y2		Ratio of volatile matter to ash for sludge.
	YTONS		Number of tons of sludge incinerated.
	ECF(I)		Excess Capacity Factor for the Ith Process. ECF is the ratio of capacity provided to that required by design relationships at the design point.
	CTRP	0.10	Contractor's profit expressed as a fraction of the total capital cost.
	CLAND	0.02	Cost of land expressed as a fraction of the total capital cost.
	CTGO	0.15	Cost of contingencies and omissions expressed as a fraction of the total capital cost.
	CENG		Cost of engineering services expressed as a fraction of the total capital cost.
	CCR		Capital Cost Ratio CCR = 1 + CENG + CTRP + CLAND + CTGO

Table A-1 VARIABLES AND PARAMETERS USED IN COST-PERFORMANCE COMPUTATION (Cont.)

Text Symbols	Computer Symbols	Nominal Value	Definitions and Comments
	TACOST		Total Amortization Cost for the Plant expressed as dollars/yr.
	TCOSTO		Total Operating and Maintenance Cost for the Plant expressed in dollars/yr.
	CAPKG		Amortization Cost, cents/1000 gallons for entire plant.
	COPKG		Operating and Maintenance Cost, cents/1000 gallons for entire plant.
	TCOST		Sum of CAPKG and COPKG.

[a] mg/l = Milligrams per liter.

PROGRAM LISTING

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C     ROBERT SMITH  SEC  COLUMBIA PARKWAY RM. 139
      DIMENSIONQ(20),SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
     1DNBC(20),DN(20),DP(20),DFM(20),SBOD(20),DBOD(20),CCOST(15),
     2COSTO(15),ACOST(15),AOCOST(15),NFORK(10),ASS(12),DEMBUD(10),
     3VSS(20),TSS(20),ALK(20),CPERKG(15),ECF(15),Q20(10),FRPSIN(10)
      DO 300 I=1,20
      Q(I)=0.0
      SOC(I)=0.0
      SNBC(I)=0.0
      SON(I)=0.0
      SOP(I)=0.0
      SFM(I)=0.0
      DOC(I)=0.0
      DNBC(I)=0.0
      DN(I)=0.0
      DP(I)=0.0
      DFM(I)=0.0
      SBOD(I)=0.0
      DBOD(I)=0.0
      VSS(I)=0.0
      TSS(I)=0.0
  300 ALK(I)=0.0
      DO 400 I=1,15
      CCOST(I)=0.0
      COSTO(I)=0.0
      ACOST(I)=0.0
      AOCOST(I)=0.0
  400 CPERKG(I)=0.0
      DO 500 I=1,12
  500 ASS(I)=0.0
      DO 600 I=1,10
      NFORK(I)=0.0
  600 DEMBOD(I)=0.0
  101 FORMAT(10F8.4/2F8.4)
  102 FORMAT(4F8.4/6F10.5/5F10.4/5F10.4/5F10.4)
  103 FORMAT(F8.2)
  107 FORMAT(I3/(F8.2))
  301 FORMAT(1H1/1X,8HCASE NO.,I4,29X,
     136HACTIVATED SLUDGE PROCESS CALCULATION,22X,9HJUNE 1967/)
  302 FORMAT(1X,7HSTATION,6X,3HMGD,7X,6HCARBON,7X,3HBOD,6X,
     110HNBIOCARBON,3X,8HNITROGEN,3X,10HPHOSPHORUS,1X,12HFIXED MATTER,
     25X,3HVSS,9X,3HTSS/)
  303 FORMAT(1X,3HSOL,I4,7E12.4)
  304 FORMAT(1X,3HDIS,16X,6E12.4/)
  105 FORMAT(7X,8HMLASS = E10.4,4X,8HMLBSS = E10.4,3X,
     19HMLNBSS = E10.4,4X,8HMLDSS = E10.4,4X,8HMLISS = E10.4/
     28X,7HMLSS = E10.4,5X,7HVAER = E10.4,5X,7HVNIT = E10.4,
     33X,9HRETURN = E10.4,6X,6HCWR = E10.4/
     46X,9HCARREM = E10.4,3X,9HBODREM = E10.4,3X,9HNITREM = E10.4,
     52X,10HPHOSREM = E10.4,6X,6HEFF = E10.4)
  111 FORMAT(8X,7HFRPS = E10.4,5X,7HURPS = E10.4,5X,7HURSS = E10.4,
   15X,7HXRSS = E10.4/9X,6HGPS = E10.4,6X,6HAPS = E10.4,
   26X,6HGSS = E10.4,6X,6HAFS = E10.4/9X,6HGTH = E10.4,
   35X,7HGSTH = E10.4,5X,7HATHM = E10.4,6X,6HTRR = E10.4/
   410X,5HGE = E10.4,6X,6HGES = E10.4,7X,5HAE = E10.4,
   56X,6HERR = E10.4,6X,6HWRE = E10.4)
112 FORMAT(8X,7HTDIG = E10.4,7X,5HTD = E10.4,5X,7HVDIG = E10.4,
   14X,8HFRDIG = E10.4/7X,8HC1DIG = E10.4,4X,8HC2DIG = E10.4,
   23X,9HCH4CFD = E10.4,3X,9HCO2CFD = E10.4/9X,6HVFL = E10.4,
   36X,6HTVF = E10.4,6X,6HAVF = E10.4,5X,7HFOOD = E10.4)
113 FORMAT(8X,7HDEGC = E10.4,7X,5HDO = E10.4,
   13X,9HCAER20 = E10.4,5X,7HCAER = E10.4/
   225X,3X,9HAEFF20 = E10.4,5X,7HAEFF = E10.4/
   325X,5X,7HCNIT = E10.4,5X,7HCEDR = E10.4/
   46X,9HAIRCFD = E10.4,3X,9HCFPGAL = E10.4,4X,8HBSIZE = E10.4)
108 FORMAT(2X,17HNAME OF STRUCTURE,8X,12HCAPITAL COST,4X,
   112HAMORTIZATION,4X,14HOPERATING COST,4X,
   232HAMORTIZATION PLUS OPERATING COST/30X,7HDOLLARS,6X,
   312HDOLLARS/YEAR,5X,12HDOLLARS/YEAR,6X,12HDOLLARS/YEAR,
   44X,14HCENTS/1000 GAL/)
114 FORMAT(10I1)
109 FORMAT(2X,21HPRELIMINARY TREATMENT,5X,E10.4,6X,E10.4,
   17X,E10.4,8X,E10.4,7X,E10.4/2X,15HPRIMARY SETTLER,11X,E10.4,6X,
   2E10.4,7X,E10.4,8X,E10.4,7X,E10.4,/2X,7HAERATOR,19X,E10.4,6X,E10.4,
   37X,E10.4,8X,E10.4,7X,E10.4/2X,11HAIR BLOWERS,15X,E10.4,6X,E10.4,
   X7X,E10.4,8X,E10.4,7X,E10.4/2X,13HFINAL SETTLER,13X,E10.4,6X,E10.4,
   X7X,E10.4,8X,E10.4,7X,E10.4/2X,19HSLUDGE RETURN PUMPS,
   X7X,E10.4,6X,E10.4,
   47X,E10.4,8X,E10.4,7X,E10.4/2X,13HCONTROL HOUSE,13X,E10.4,6X,E10.4,
   57X,E10.4,8X,E10.4,7X,E10.4)
118 FORMAT(7X,8HFECL3 = E10.4,3X,9HALK(1) = E10.4,2X,
   110HALK(12) = E10.4,2X,10HALK(13) = E10.4,2X,10HALK(15) = E10.4)
119 FORMAT(2X,16HSLUDGE THICKENER,10X,E10.4,6X,E10.4,
   17X,E10.4,8X,E10.4,7X,E10.4/2X,8HDIGESTER,18X,E10.4,6X,
   2E10.4,7X,E10.4,8X,E10.4,7X,E10.4/2X,18HSLUDGE ELUTRIATION,
   X8X,E10.4,
   36X,E10.4,7X,E10.4,8X,E10.4,7X,E10.4/2X,17HVACUUM FILTRATION,
   X9X,E10.4,
   46X,E10.4,7X,E10.4,8X,E10.4,7X,E10.4/2X,19HSLUDGE INCINERATION,
   X7X,E10.4,
   56X,E10.4,7X,E10.4,8X,E10.4,7X,E10.4/2X,18HSLUDGE DRYING BEDS,
   68X,E10.4,6X,E10.4,7X,E10.4,8X,E10.4,7X,E10.4)
110 FORMAT(2X,18HTOTAL CAPITAL COST,8X,E10.4/2X,
   137HTOTAL AMORTIZATION AND OPERATING COST,49X,E10.4/2X,
   221HCOST PER 1000 GALLONS,21X,E10.4,7X,E10.4,17X,E10.4//////)
306 FORMAT(6X,30HBOD5 DEMAND CANNOT BE ACHIEVED)
115 FORMAT(25X,4HMG/L,8X,4HMG/L,8X,4HMG/L,8X,4HMG/L,8X,
   14HMG/L,8X,4HMG/L,9X,4HMG/L,8X,4HMG/L)
116 FORMAT(10X,11HPERCENT H20,3X,6HLB/DAY,5X,6HLB/DAY,7X,
   16HLB/DAY,6X,6HLB/DAY,6X,6HLB/DAY,6X,6HLB/DAY,7X,6HLB/DAY,
   26X,6HLB/DAY)
307 FORMAT(6X,27HDEMAND MLASS CANNOT BE HELD)
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  313 FORMAT(1X,3HSOL,I4,9E12.4)
  308 FORMAT(6X,5HECF  ,15F5.2)
  309 FORMAT(2X,12HCHLORINATION,14X,E10.4,6X,E10.4,7X,E10.4,
     18X,E10.4,7X,E10.4/2X,16HSITE DEVELOPMENT,10X,E10.4,6X,E10.4,
     27X,E10.4,8X,E10.4,7X,E10.4/)
  305 FORMAT (15F4.2)
  700 FORMAT(6X,7HCENG = E10.4,2X,7HCTRP = E10.4,2X,7HCTGO = E10.4,
     12X,8HCLAND = E10.4,2X,6HCCR = E10.4)
  701 FORMAT(6X,6HCCI = E10.4,6X,5HAF = E10.4)
      READ(5,305) (ECF(I),I=1,15)
      READ(5,114) (NFORK(I),I=1,10)
      READ(5,101) SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
     1DNBC(20),DN(20),DP(20),DFM(20),DEGC,ALK(20)
      VSS(20)=SOC(20)*2.13
      TSS(20)=VSS(20)+SFM(20)
      SBOD(20)=(SOC(20)-SNBC(20))*1.87
      DBOD(20)=(DOC(20)-DNBC(20))*1.87
      READ(5,102) URPS,URSS,XRSS,CAER20,AEFF20,DO,CKWH,CCI,AF,GSS,
     1TRR,TSS(12),GTH,GSTH,ERR,TSS(15),WRE,GE,GES,TDIG,TD,
     2TSS(16),VFL,TVF,SBL
      AF=.06744
      CCI=1.13
      CCL2=.08
      DCL2=8.
      READ(5,107) NAS,(ASS(I),I=1,NAS)
      READ(5,107) NBOD,(DEMBOD(J),J=1,NBOD)
      READ(5,107)NQ,(Q20(J),J=1,NQ)
      READ(5,107)NFR,(FRPSIN(J),J=1,NFR)
      LOOPS=6
      DO 25 LF=1,NFR
      FRPS=FRPSIN(LF)
      DO 25 LQ=1,NQ
      Q(20)=Q20(LQ)
   80 DO25I=1,NAS
      XMLSS=ASS(I)
      DO25K=1,NBOD
      BOD5=DEMBOD(K)
   67 IF(LOOPS-6)66,65,66
C     MIX STREAMS NINE AND TWENTY
   65 Q(9)=0.0
   66 TEMP1=Q(20)/(Q(20)+Q(9))
      TEMP2=Q(9)/(Q(20)+Q(9))
      SOC(1)=TEMP1*SOC(20)+TEMP2*SOC(9)
      SON(1)=TEMP1*SON(20)+TEMP2*SON(9)
      SOP(1)=TEMP1*SOP(20)+TEMP2*SOP(9)
      SNBC(1)=TEMP1*SNBC(20)+TEMP2*SNBC(9)
      SFM(1)=TEMP1*SFM(20)+TEMP2*SFM(9)
      DOC(1)=TEMP1*DOC(20)+TEMP2*DOC(9)
      DNBC(1)=TEMP1*DNBC(20)+TEMP2*DNBC(9)
      DN(1)=TEMP1*DN(20)+TEMP2*DN(9)
      DP(1)=TEMP1*DP(20)+TEMP2*DP(9)
      DFM(1)=TEMP1*DFM(20)+TEMP2*DFM(9)
      ALK(1)=TEMP1*ALK(20)+TEMP2*ALK(9)
      Q(1)=Q(20)+Q(9)
      VSS(1)=SOC(1)*2.13
      TSS(1)=VSS(1)+SFM(1)
      SBOD(1)=(SOC(1)-SNBC(1))*1.87
      DBOD(1)=(DOC(1)-DNBC(1))*1.87
C     PRIMARY SETTLER PERFORMANCE
      Q(8)=FRPS*Q(1)/URPS
      Q(2)=Q(1)-Q(8)
      TEMP1=(1.-FRPS)*Q(1)/Q(2)
      SOC(2)=TEMP1*SOC(1)
      SNBC(2)=TEMP1*SNBC(1)
      SON(2)=TEMP1*SON(1)
      SOP(2)=TEMP1*SOP(1)
      SFM(2)=TEMP1*SFM(1)
      DOC(2)=DOC(1)
      DNBC(2)=DNBC(1)
      DN(2)=DN(1)
      DP(2)=DP(1)
      DFM(2)=DFM(1)
      VSS(2)=SOC(2)*2.13
      TSS(2)=VSS(2)+SFM(2)
      SBOD(2)=(SOC(2)-SNBC(2))*1.87
      DBOD(2)=(DOC(2)-DNBC(2))*1.87
      BOD2=SBOD(2)+DBOD(2)
      DBOD2=DBOD(2)
      TEMP1=FRPS*Q(1)/Q(8)
      SOC(8)=TEMP1*SOC(1)
      SNBC(8)=TEMP1*SNBC(1)
      SON(8)=TEMP1*SON(1)
      SOP(8)=TEMP1*SOP(1)
      SFM(8)=TEMP1*SFM(1)
      DOC(8)=DOC(2)
      DNBC(8)=DNBC(2)
      DN(8)=DN(2)
      DP(8)=DP(2)
      DFM(8)=DFM(2)
      VSS(8)=SOC(8)*2.13
      TSS(8)=VSS(8)+SFM(8)
      SBOD(8)=(SOC(8)-SNBC(8))*1.87
      DBOD(8)=(DOC(8)-DNBC(8))*1.87
C     AERATOR PERFORMANCE
 1000 CEDR=.18*1.047**(DEGC-28.)
      CAER=CAER20*1.047**(DEGC-20.)
      SVI=100.
      SMAX=1200000./SVI
      IF (URSS*XMLSS-SMAX) 86,86,85
   85 URSS=SMAX/XMLSS
   86 SA=XMLSS/1000.
      TA=(BOD2-BOD5)/(BOD5*CAER*SA*24.)
      VAER=Q(2)*TA
      XRSS=556.1*GSS**.4942/XMLSS**1.8165/(TA*24.)**.4386
      ASMAX=BOD5/XRSS/.685
      ASMIN=0.0
      IF(ASMAX-XMLSS)50,50,70
   70 ASMAX=XMLSS
   50 XMLASS=(ASMAX+ASMIN)/2.0
   42 FOOD=SBOD(2)+DBOD2
      FMAX=FOOD
      N=1
      GO TO 8
    7 ERROR=FMAX-FMIN
      TOL=.10
      IF(ERROR-TOL)21,21,19
   19 FOOD=(FMIN+FMAX)/2.0
    4 IF(FOOD-DBOD(2))5,5,6
    5 DBOD(4)=DBOD(2)-FOOD
      SBOD(4)=SBOD(2)
      GO TO 8
    6 SBOD(4)=(SBOD(2)+DBOD(2)-FOOD)*.70
      DBOD(4)=.233*SBOD(4)
    8 TEMP1=(.65*FOOD/XMLASS)-XRSS
      Q(7)=(Q(2)*TEMP1-CEDR*VAER)/(URSS-XRSS)
   20 Q(5)=Q(2)-Q(7)
      N=N+1
      IF(N-2)23,23,22
   22 TEMP2=XRSS*Q(5)+URSS*Q(7)
      XMLBSS=Q(2)*SBOD(4)/TEMP2/.80
      SBOD(5)=(XMLASS*.685+XMLBSS*.80)*XRSS
      DBOD(5)=DBOD(4)
      TBOD5=SBOD(5)+DBOD(5)
      GO TO 24
   23 TBOD5=XMLASS*XRSS*.685
   24 IF(TBOD5-BOD5)10,10,15
   10 IF(N-3)11,12,13
   11 FMIN=(CEDR*VAER/Q(2)+XRSS)*XMLASS/.65
      FOOD=FMIN
      GO TO 4
   12 WRITE(6,307)
      GO TO 25
   13 FMAX=FOOD
      GO TO 7
   15 IF(N-3)16,18,17
   16 WRITE(6,306)
      GO TO 25
   17 FMIN=FOOD
   18 GO TO 7
   21 CONTINUE
      XLNBSS=SNBC(2)*Q(2)*2.13/TEMP2
      XMLISS=Q(2)*SFM(2)/TEMP2
      XMLDSS=(.12*Q(2)*FOOD/TEMP2)-.185*XMLASS
      TEMP1=XMLASS+XMLBSS+XLNBSS+XMLDSS+XMLISS
      TEMP2=TEMP1-XMLSS
      TOL=5.0
      IF(ABS(TEMP2)-TOL)41,41,51
   51 IF(TEMP2)52,52,53
   52 ASMIN=XMLASS
      GO TO 50
   53 ASMAX=XMLASS
      GO TO 50
   41 CONTINUE
      SOC(5)=(XMLDSS+XMLASS)*XRSS/2.46+(XMLBSS+XLNBSS)*XRSS/2.33
      SOC(7)=SOC(5)*URSS/XRSS
      SNBC(5)=XLNBSS*XRSS/2.33+(XMLDSS+.185*XMLASS)*XRSS/2.46
      SNBC(7)=SNBC(5)*URSS/XRSS
      TEMP2=XRSS*XMLASS/2.46
      SON(5)=.234*TEMP2+(SOC(5)-TEMP2)/10.
      TEMP2=TEMP2*URSS/XRSS
      SON(7)=.234*TEMP2+(SON(7)-TEMP2)/10.
      SOP(5)=SOC(5)*.01
      SOP(7)=SOC(7)*.01
      SFM(5)=XMLISS*XRSS
      SFM(7)=XMLISS*URSS
      DOC(5)=DNBC(2)+DBOD(4)/1.87
      DNBC(5)=DNBC(2)
      DOC(7)=DOC(5)
      DNBC(7)=DNBC(5)
      DN(5)=(Q(2)*(SON(2)+DN(2))-(SON(5)*Q(5)+SON(7)*Q(7)))/(Q(5)+Q(7))
      DN(7)=DN(5)
      DP(5)=(Q(2)*(SOP(2)+DP(2))-(SOP(5)*Q(5)+SOP(7)*Q(7)))/(Q(5)+Q(7))
      DP(7)=DP(5)
      DFM(5)=DFM(2)
      DFM(7) = DFM(2)
      SBOD(7)=(SOC(7)-SNBC(7))*1.87
      DBOD(7)=(DOC(7)-DNBC(7))*1.87
      VSS(5)=SOC(5)*1.90
      TSS(5)=VSS(5)+SFM(5)
      VSS(7)=SOC(7)*1.90
      TSS(7)=VSS(7)+SFM(7)
C     CONDITIONS FOR NITRIFICATION
      Q(6)=(Q(2)*(1.-.65*FOOD/XMLASS)+CEDR*VAER)/(URSS-1.)
      RETURN=Q(6)/Q(2)
      X4X3=(1.+RETURN)/RETURN/URSS
      DN(3)=DN(2)/(1.+RETURN)
      X3Y=DN(3)*.99/(X4X3-1.)
      CNIT=.18*EXP(.116*(DEGC-15.))
      TAN=(1.+RETURN)*(ALOG(X4X3)+4.605/(DN(3)+X3Y))/CNIT
      VNIT=Q(2)*TAN
C     AIR REQUIREMENTS
      DOSAT=14.16-.3943*DEGC+.007714*DEGC**2-.0000646*DEGC**3
      DOSAT=DOSAT*1.221
      AEFF=AEFF20*(DOSAT-DO)*1.02**(DEGC-20.)/DOSAT
      WFOOD=Q(2)*FOOD*8.33
      WAS=XMLASS*VAER*8.33
      AIRCFD=(.577*WFOOD+1.16*CEDR*WAS)/AEFF/.232/.075
      WDN=(Q(5)*DN(5)+Q(7)*DN(7))*8.33
      IF(VNIT-VAER)26,26,27
   26 AIRCFD=AIRCFD+4.6*WDN/AEFF/.232/.075
   27 BSIZE=AIRCFD/1440.
      CFPGAL=AIRCFD/1000000./Q(1)
C     CALCULATE STREAM INTO THICKENER
      Q(10)=Q(7)+Q(8)
      SOC(10)=(SOC(7)*Q(7)+SOC(8)*Q(8))/Q(10)
      SNBC(10)=(SNBC(7)*Q(7)+SNBC(8)*Q(8))/Q(10)
      SON(10)=(SON(7)*Q(7)+SON(8)*Q(8))/Q(10)
      SOP(10)=(SOP(7)*Q(7)+SOP(8)*Q(8))/Q(10)
      SFM(10)=(SFM(7)*Q(7)+SFM(8)*Q(8))/Q(10)
      DOC(10)=(DOC(7)*Q(7)+DOC(8)*Q(8))/Q(10)
      DNBC(10)=(DNBC(7)*Q(7)+DNBC(8)*Q(8))/Q(10)
      DN(10)=(DN(7)*Q(7)+DN(8)*Q(8))/Q(10)
      DP(10)=(DP(7)*Q(7)+DP(8)*Q(8))/Q(10)
      DFM(10)=(DFM(7)*Q(7)+DFM(8)*Q(8))/Q(10)
      VSS(10)=SOC(10)*2.0
      TSS(10)=VSS(10)+SFM(10)
C     CALCULATE STREAMS FROM THICKENER
      Q(12)=TRR*Q(10)*TSS(10)/TSS(12)
      Q(11)=Q(10)-Q(12)
      TSS(11)=(1.-TRR)*Q(10)*TSS(10)/Q(11)
      TEMP4=TSS(11)/TSS(10)
      SOC(11)=TEMP4*SOC(10)
      VSS(11)=SOC(11)*2.0
      SNBC(11)=TEMP4*SNBC(10)
      SON(11)=TEMP4*SON(10)
      SOP(11)=TEMP4*SOP(10)
      SFM(11)=TEMP4*SFM(10)
      DOC(11)=DOC(10)
      DNBC(11)=DNBC(10)
      DN(11)=DN(10)
      DP(11)=DP(10)
      DFM(11)=DFM(10)
      TEMP3=TSS(12)/TSS(10)
      ATH2=Q(10)*TSS(10)*8.33/GSTH
      SOC(12)=TEMP3*SOC(10)
      SNBC(12)=TEMP3*SNBC(10)
      SON(12)=TEMP3*SON(10)
      SOP(12)=TEMP3*SOP(10)
      SFM(12)=TEMP3*SFM(10)
      DOC(12)=DOC(10)
      DNBC(12)=DNBC(10)
      DN(12)=DN(10)
      DP(12)=DP(10)
      DFM(12)=DFM(10)
      VSS(12)=SOC(12)*2.0
      ATH1=Q(10)*1000000./GTH
      ATHM=AMAX1(ATH1,ATH2)
C     CALCULATE DIGESTER PERFORMANCE
      C1DIG=.28/EXP(.036*(35.-TDIG))
      C2DIG=700.*EXP(.10*(35.-TDIG))
      DIGC12=SOC(12)-SNBC(12)+DOC(12)-DNBC(12)
      TDM=2.5/C1DIG
      TDOPT=(1.-(C2DIG/(C2DIG+DIGC12)**.5))/C1DIG
      TDMIN=AMAX1(TDM,TDOPT)
      IF(TD-TDMIN) 43,43,44
   43 TD=TDMIN
   44 DIGC13=C2DIG/(C1DIG*TD-1.)
      TEMP4=(DIGC12-DIGC13)/(SOC(12)+DOC(12))
      CDF=200.*EXP(.12*(35.-TDIG))
      DF=CDF/(C1DIG*TD-1.)
      SOC(13)=SNBC(12)+DIGC13-DF
      DOC(13)=DNBC(12)+DF
      FRDIG=(SOC(12)-SOC(13))/SOC(12)
      SNBC(13)=SNBC(12)
      SON(13)=(1.-TEMP4)*SON(12)
      SOP(13)=(1.-TEMP4)*SOP(12)
      SFM(13)=SFM(12)
      DNBC(13)=DNBC(12)
      DN(13)=DN(12)+.65*SON(12)*TEMP4
      DP(13)=DP(12)+SOP(12)*TEMP4
      DFM(13)=DFM(12)
      VDIG=Q(12)*TD*1000./7.48
      CH4CFD=163.85*(DIGC12-DIGC13)*Q(12)
      CO2CFD=249.9*(DIGC12-DIGC13)*Q(12)-CH4CFD
      VSS(13)=SOC(13)*2.0
      TSS(13)=VSS(13)+SFM(13)
      Q(13)=Q(12)
C     CALCULATE STREAMS FROM SLUDGE WASH
      IF(NFORK(1))45,45,46
   45 Q(17)=WRE*Q(13)
      TEMP1=Q(13)*1000000./GE
      TEMP2=Q(13)*TSS(13)*8.33/GES
      AE=AMAX1(TEMP1,TEMP2)
      Q(15)=ERR*Q(13)*TSS(13)/TSS(15)
      Q(14)=Q(17)+Q(13)-Q(15)
      TSS(14)=Q(13)*((1.-ERR)*TSS(13)+WRE*TSS(5))/Q(14)
      VSS(17)=VSS(5)
      TSS(17)=TSS(5)
      TEMP1=TSS(15)/TSS(13)
      SOC(15)=TEMP1*SOC(13)
      SNBC(15)=TEMP1*SNBC(13)
      SON(15)=TEMP1*SON(13)
      SOP(15)=TEMP1*SOP(13)
      SFM(15)=TEMP1*SFM(13)
      VSS(15)=SOC(15)*2.0
      TEMP2=TSS(14)/TSS(13)
      SOC(14)=TEMP2*SOC(13)
      VSS(14)=SOC(14)*2.0
      SNBC(14)=TEMP2*SNBC(13)
      SON(14)=TEMP2*SON(13)
      SOP(14)=TEMP2*SOP(13)
      SFM(14)=TEMP2*SFM(13)
      TEMP1=Q(13)/(Q(13)+Q(17))
      TEMP2=Q(17)/(Q(13)+Q(17))
      DOC(14)=TEMP1*DOC(13)+TEMP2*DOC(5)
      DNBC(14)=TEMP1*DNBC(13)+TEMP2*DNBC(5)
      DN(14)=TEMP1*DN(13)+TEMP2*DN(5)
      DP(14)=TEMP1*DP(13)+TEMP2*DP(5)
      DFM(14)=TEMP1*DFM(13)+TEMP2*DFM(5)
      DOC(15)=DOC(14)
      DNBC(15)=DNBC(14)
      DN(15)=DN(14)
      DP(15)=DP(14)
      DFM(15)=DFM(14)
      IF(VNIT-VAER)60,60,61
   60 ALK(5)=ALK(1)+3.57*(DN(5)-DN(2))
      GO TO 62
   61 ALK(5)=ALK(1)
   62 ALK(12)=ALK(5)
      ALK(13)=ALK(12)+(DN(13)-DN(12))*3.57
      ALK(14)=TEMP1*ALK(13)+TEMP2*ALK(5)
      ALK(15)=ALK(14)
      Y1=100.*ALK(15)/TSS(15)
      Y2=VSS(15)/(TSS(15)-VSS(15))
      FECL3=1.08*Y1+2.*Y2
      CFECL3=.08
C     CALCULATE STREAMS FROM VACUUM FILTER
      WP=88./(TSS(15)/10000.)**.123
      TEMP1=1000000.*(100.-WP)/WP
      Q(16)=Q(15)*(TEMP1-TSS(15))/(TEMP1-TSS(16))
      TSS(18)=8.33*(Q(15)*TSS(15)-Q(16)*TSS(16))
      TEMP2=TSS(18)/TSS(15)
      SOC(18)=TEMP2*SOC(15)
      VSS(18)=SOC(18)*2.0
      SNBC(18)=TEMP2*SNBC(15)
      SON(18)=TEMP2*SON(15)
      SOP(18)=TEMP2*SOP(15)
      VSS(16)=SOC(16)*2.0
      DOC(16)=DOC(15)
      DNBC(16)=DNBC(15)
      DN(16)=DN(15)
      DP(16)=DP(15)
      TEMP3=8.33*(Q(15)-Q(16))
      DFM(18)=DFM(15)*TEMP3
      DP(18)=DP(15)*TEMP3
      DN(18)=DN(15)*TEMP3
      DNBC(18)=DNBC(15)*TEMP3
      DOC(18)=DOC(15)*TEMP3
      TEMP2=TSS(16)/TSS(15)
      SOC(16)=SOC(15)*TEMP2
      SNBC(16)=SNBC(15)*TEMP2
      SON(16)=SON(15)*TEMP2
      SOP(16)=SOP(15)*TEMP2
      SFM(16)=SFM(15)*TEMP2
      DFM(16)=DFM(15)
      AVF=Q(16)*1000000./VFL/TVF/24.
      Q(9)=Q(11)+Q(14)+Q(16)
      TEMP1=Q(11)/Q(9)
      TEMP2=Q(14)/Q(9)
      TEMP3=Q(16)/Q(9)
      SOC(9)=TEMP1*SOC(11)+TEMP2*SOC(14)+TEMP3*SOC(16)
      SNBC(9)=TEMP1*SNBC(11)+TEMP2*SNBC(14)+TEMP3*SNBC(16)
      SON(9)=TEMP1*SON(11)+TEMP2*SON(14)+TEMP3*SON(16)
      SOP(9)=TEMP1*SOP(11)+TEMP2*SOP(14)+TEMP3*SOP(16)
      SFM(9)=TEMP1*SFM(11)+TEMP2*SFM(14)+TEMP3*SFM(16)
      DOC(9)=TEMP1*DOC(11)+TEMP2*DOC(14)+TEMP3*DOC(16)
      DNBC(9)=TEMP1*DNBC(11)+TEMP2*DNBC(14)+TEMP3*DNBC(16)
      DN(9)=TEMP1*DN(11)+TEMP2*DN(14)+TEMP3*DN(16)
      DP(9)=TEMP1*DP(11)+TEMP2*DP(14)+TEMP3*DP(16)
      DFM(9)=TEMP1*DFM(11)+TEMP2*DFM(14)+TEMP3*DFM(16)
      SBOD(9)=(SOC(9)-SNBC(9))*1.87
      DBOD(9)=(DOC(9)-DNBC(9))*1.87
      SBOD(9)=(SOC(9)-SNBC(9))*1.87
      VSS(9)=SOC(9)*2.0
      TSS(9)=VSS(9)+SFM(9)
      Q(4)=Q(2)+Q(6)
C     CALCULATE CAPITAL AND OPERATING COST
   46 CCOST(1)=14700.*Q(20)**.625
      COSTO(1)=500.*Q(20)+2150.*Q(20)**.37
      GPS=-2780.*ALOG(FRPS)-551.7
      APS=Q(1)*1000./GPS
      APS=APS*ECF(2)
      CCOST(2)=13400.*APS+5200.*APS**.10
      COSTO(2)=1000.*APS+2500.*APS**.5
 1001 VAER=VAER*ECF(3)
      CCOST(3)=175000.*VAER+36500.*VAER**.182
      COSTO(3)=10000.*VAER+14500.*VAER**.5
      BSIZE=BSIZE*ECF(4)
      CCOST(4)=10570.+5.857*BSIZE
      CAIRP=AIRCFD*365.*CKWH/1830.
      COSTO(4)=CAIRP
      AFS=Q(4)*1000./GSS
      AFS=AFS*ECF(5)
      CCOST(5)=12600.*AFS+5350./AFS**.126
      Q(6)=Q(6)*ECF(6)
      CCOST(6)=3650.+2250.*Q(6)
      CCOST(7)=40000.*Q(20)**.70
   ATHM=ATHM*ECF(8)
   CCOST(8)=(18.8+9.1/EXP(ATHM/13300.))*ATHM
   VDIG=VDIG*ECF(9)
   CCOST(9)=5000.+1080.*VDIG+10700.*VDIG**.128
   AE=AE*ECF(10)
   CCOST(10)=(18.8+52.0/EXP(AE/6000.))*AE
   AVF=AVF*ECF(11)
   CCOST(11)=12800.+372.*AVF
   COSTO(11)=TSS(15)*FECL3*Q(15)*30.40*CFECL3
   COSTO(11)=COSTO(11)+1500.*Q(20)+6450.*Q(20)**.37
   YTONS=.1825*TSS(18)
   COSTO(12)=16.1*YTONS-.00009*YTONS**2.
   CCOST(12)=1570.*TSS(18)**.60
   CCOST(14)=9000.*(Q(5)-Q(17))**.469
   COSTO(14)=DCL2*(Q(5)-Q(17))*8.33*CCL2*365.
   CCOST(15)=4400.*Q(20)**.875
   IF(NFORK(1))48,48,47
47 ASB=TSS(13)*Q(13)*8.33*30./SBL
   CCOST(10)=0.0
   CCOST(11)=0.0
   CCOST(13)=2.23*ASB
   COSTO(9)=80.*VDIG+900.*VDIG**.44
   GO TO 49
48 COSTO(9)=48.*VDIG+540.*VDIG**.44
   CTRP=.10
   CTGO=.15
   CLAND=.02
49 DO30J=1,15
30 CCOST(J)=CCOST(J)*CCI
   DO31J=1,15
31 ACOST(J)=CCOST(J)*AF
   TOTCC=0.0
   DO32J=1,15
32 TOTCC=TOTCC+CCOST(J)
   CENG=.08*(1000000./TOTCC)**.146
   CCR=1.+CENG+CTRP+CTGO+CLAND
   TOTCC=TOTCC*CCR
   DO90J=1,15
   ACOST(J)=ACOST(J)*CCR
90 CCOST(J)=CCOST(J)*CCR
   TACOST=0.0
   TCOSTO=0.0
   TOTAOC=0.0
   DO 33 J=1,15
   TACOST=TACOST+ACOST(J)
   AOCOST(J)=ACOST(J)+COSTO(J)
   CPERKG(J)=AOCOST(J)/Q(20)/3650.
   TOTAOC=TOTAOC+AOCOST(J)
33 TCOSTO=TCOSTO+COSTO(J)
   CAPKG=TACOST/Q(20)/3650.
   COPKG=TCOSTO/Q(20)/3650.
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      TCOST=TOTAOC/Q(20)/3650.
      TOC2=SOC(2)+DOC(2)
      TOC5=SOC(5)+DOC(5)
      TOC7=SOC(7)+DOC(7)
      TBOD2=SBOD(2)+DBOD(2)
      TBOD5=SBOD(5)+DBOD(5)
      CWR=Q(7)*TOC7/Q(2)/TOC2
      CREM=1.-TOC5*Q(5)/Q(2)/TOC2
      BODREM=1.-TBOD5*Q(5)/Q(2)/TBOD2
      EFF=CREM-CWR
      REMN=1.-(SON(5)+DN(5))*Q(5)/Q(2)/(SON(2)+DN(2))
      REMP=1.-(SOP(5)+DP(5))*Q(5)/Q(2)/(SOP(2)+DP(2))
      LOOPS=LOOPS-1
      IF(LOOPS)67,68,67
   68 NCASE=1
      WRITE(6,301) NCASE
      WRITE (6,115)
      WRITE(6,302)
      NSTA=1
      WRITE(6,313) NSTA,Q(1),SOC(1),SBOD(1),SNBC(1),SON(1),SOP(1),
     1SFM(1),VSS(1),TSS(1)
      WRITE(6,304) DOC(1),DBOD(1),DNBC(1),DN(1),DP(1),DFM(1)
      NSTA=2
      WRITE(6,313) NSTA,Q(2),SOC(2),SBOD(2),SNBC(2),SON(2),SOP(2),
     1SFM(2),VSS(2),TSS(2)
      WRITE(6,304) DOC(2),DBOD(2),DNBC(2),DN(2),DP(2),DFM(2)
      NSTA=8
      WRITE(6,313) NSTA,Q(8),SOC(8),SBOD(8),SNBC(8),SON(8),SOP(8),
     1SFM(8),VSS(8),TSS(8)
      WRITE(6,304) DOC(8),DBOD(8),DNBC(8),DN(8),DP(8),DFM(8)
      NSTA=5
      WRITE(6,313) NSTA,Q(5),SOC(5),SBOD(5),SNBC(5),SON(5),SOP(5),
     1SFM(5),VSS(5),TSS(5)
      WRITE(6,304) DOC(5),DBOD(5),DNBC(5),DN(5),DP(5),DFM(5)
      NSTA=7
      WRITE(6,313) NSTA,Q(7),SOC(7),SBOD(7),SNBC(7),SON(7),SOP(7),
     1SFM(7),VSS(7),TSS(7)
      WRITE(6,304) DOC(7),DBOD(7),DNBC(7),DN(7),DP(7),DFM(7)
      NSTA=10
      WRITE(6,313) NSTA,Q(10),SOC(10),SBOD(10),SNBC(10),SON(10),
     1SOP(10),SFM(10),VSS(10),TSS(10)
      WRITE(6,304) DOC(10),DBOD(10),DNBC(10),DN(10),DP(10),DFM(10)
      NSTA=11
      WRITE(6,313) NSTA,Q(11),SOC(11),SBOD(11),SNBC(11),SON(11),
     1SOP(11),SFM(11),VSS(11),TSS(11)
      WRITE(6,304) DOC(11),DBOD(11),DNBC(11),DN(11),DP(11),DFM(11)
      NSTA=12
      WRITE(6,313) NSTA,Q(12),SOC(12),SBOD(12),SNBC(12),SON(12),SOP(12),
     1SFM(12),VSS(12),TSS(12)
      WRITE(6,304) DOC(12),DBOD(12),DNBC(12),DN(12),DP(12),DFM(12)
      NSTA=13
      WRITE(6,313) NSTA,Q(13),SOC(13),SBOD(13),SNBC(13),SON(13),
     1SOP(13),SFM(13),VSS(13),TSS(13)
      WRITE(6,304) DOC(13),DBOD(13),DNBC(13),DN(13),DP(13),DFM(13)
      NSTA=14
      WRITE(6,313) NSTA,Q(14),SOC(14),SBOD(14),SNBC(14),SON(14),
     1SOP(14),SFM(14),VSS(14),TSS(14)
      WRITE(6,304) DOC(14),DBOD(14),DNBC(14),DN(14),DP(14),DFM(14)
      NSTA=15
      WRITE(6,313) NSTA,Q(15),SOC(15),SBOD(15),SNBC(15),SON(15),
     1SOP(15),SFM(15),VSS(15),TSS(15)
      WRITE(6,304) DOC(15),DBOD(15),DNBC(15),DN(15),DP(15),DFM(15)
      NSTA=16
      WRITE(6,313) NSTA,Q(16),SOC(16),SBOD(16),SNBC(16),SON(16),
     1SOP(16),SFM(16),VSS(16),TSS(16)
      WRITE(6,304) DOC(16),DBOD(16),DNBC(16),DN(16),DP(16),DFM(16)
      NSTA=9
      WRITE(6,313) NSTA,Q(9),SOC(9),SBOD(9),SNBC(9),SON(9),SOP(9),
     1SFM(9),VSS(9),TSS(9)
      WRITE(6,304) DOC(9),DBOD(9),DNBC(9),DN(9),DP(9),DFM(9)
      WRITE(6,116)
      WRITE(6,302)
      NSTA=18
      WRITE(6,313) NSTA,WP,SOC(18),SBOD(18),SNBC(18),SON(18),SOP(18),
     1SFM(18),VSS(18),TSS(18)
      WRITE(6,304) DOC(18),DBOD(18),DNBC(18),DN(18),DP(18),DFM(18)
      NSTA=20
      WRITE(6,313) NSTA,Q(20),SOC(20),SBOD(20),SNBC(20),SON(20),
     1SOP(20),SFM(20),VSS(20),TSS(20)
      WRITE(6,304) DOC(20),DBOD(20),DNBC(20),DN(20),DP(20),DFM(20)
      WRITE(6,301) NCASE
      WRITE(6,108)
      WRITE(6,109) (CCOST(L),ACOST(L),COSTO(L),AOCOST(L),
     1CPERKG(L),L=1,7)
      WRITE(6,119) (CCOST(L),ACOST(L),COSTO(L),AOCOST(L),
     1CPERKG(L),L=8,13)
      WRITE(6,309) (CCOST(L),ACOST(L),COSTO(L),AOCOST(L),
     1CPERKG(L),L=14,15)
      WRITE(6,110)TOTCC,TOTAOC,CAPKG,COPKG,TCOST
      WRITE(6,105) XMLASS,XMLBSS,XLNBSS,XMLDSS,XMLISS,XMLSS,VAER,
     1VNIT,RETURN,CWR,CREM,BODREM,REMN,REMP,EFF
      WRITE(6,111) FRPS,URPS,URSS,XRSS,GPS,APS,GSS,AFS,GTH,GSTH,ATHM,
     1TRP,GE,GES,AE,ERR,WRE
      WRITE(6,112) TDIG,TD,VDIG,FRDIG,C1DIG,C2DIG,CH4CFD,CO2CFD,VFL,
     1TVF,AVF,FOOD
      WRITE(6,113) DEGC,DO,CAER20,CAER,AEFF20,AEFF,CNIT,CEDR,
     1AIRCFD,CFPGAL,BSIZE
      WRITE(6,118) FECL3,ALK(1),ALK(12),ALK(13),ALK(15)
      WRITE(6,308)(ECF(J),J=1,15)
      WRITE(6,700)CENG,CTRP,CTGO,CLAND,CCR
      WRITE(6,701)CCI,AF
      LOOPS=6
   25 CONTINUE
```

```
      STOP
      END
      INPUT DATA


1.002.001.201.502.002.001.001.502.001.501.001.001.001.001.00
0000000000
105.0000 30.0000 10.0000  2.0000 30.0000 43.0000 11.0000 19.1000  4.0000500.000
 20.0000250.0000
400.0000  3.0000   .0200  1.0000
    .05000   1.00000    .01000   1.25000    .05783 2000.0000
     .950060000.0000  750.0000    9.0000     .7600
60000.0000    3.0000  800.0000    9.0000   33.0000
   15.0000  200.0000    4.9000     .2380    4.4000
  1
 2000.00
  1
   13.00
  1
   10.00
  1
     .50
/*
//
//
```

```
CASE NO.   1                            ACTIVATED SLUDGE PROCESS CALCULATION                         JUNE 1967

                           MG/L        MG/L          MG/L         MG/L         MG/L        MG/L          MG/L         MG/L
STATION     MGD           CARBON       BOD        NBIOCARBON    NITROGEN    PHOSPHORUS FIXED MATTER      VSS          TSS

SOL    1  0.1032E 02   0.1160E 03  0.1418E 03  0.4021E 02  0.1086E 02  0.2154E 01  0.3912E 02  0.2471E 03  0.2863E 03
DIS                    0.4241E 02  0.5874E 02  0.1100E 02  0.2210E 02  0.4860E 01  0.5000E 03

SOL    2  0.1031E 02   0.5809E 02  0.7098E 02  0.2013E 02  0.3436E 01  0.1078E 01  0.1958E 02  0.1237E 03  0.1433E 03
DIS                    0.4241E 02  0.5874E 02  0.1100E 02  0.2210E 02  0.4860E 01  0.5000E 03

SOL    8  0.1290E-01   0.4641E 05  0.5672E 05  0.1608E 05  0.4343E 04  0.8617E 03  0.1565E 05  0.9886E 05  0.1145E 06
DIS                    0.4241E 02  0.5874E 02  0.1100E 02  0.2210E 02  0.4860E 01  0.5000E 03

SOL    5  0.1011E 02   0.8903E 01  0.9285E 01  0.4261E 01  0.1401E 01  0.8903E-01  0.3532E 01  0.1692E 02  0.2045E 02
DIS                    0.1298E 02  0.3695E 01  0.1100E 02  0.2357E 02  0.5444E 01  0.5000F 03

SOL    7  0.1947E 00   0.2150E 04  0.2096E 04  0.1029E 04  0.1371E 03  0.2150E 02  0.8531E 03  0.4086E 04  0.4939E 04
DIS                    0.1298E 02  0.3695E 01  0.1100E 02  0.2357E 02  0.5444E 01  0.5000E 03

SOL   10  0.2076E 00   0.4900E 04  0.0         0.1965E 04  0.3985E 03  0.7371E 02  0.1772E 04  0.9801E 04  0.1157E 05
DIS                    0.1481E 02  0.0         0.1100E 02  0.2348E 02  0.5408E 01  0.5000E 03

SOL   11  0.1695E 00   0.3000E 03  0.0         0.1203E 03  0.2440E 02  0.4512E 01  0.1085E 03  0.6000E 03  0.7085E 03
DIS                    0.1481E 02  0.0         0.1100E 02  0.2348E 02  0.5408E 01  0.5000E 03

SOL   12  0.3804E-01   0.2541E 05  0.0         0.1018E 05  0.2066E 04  0.3821E 03  0.9188E 04  0.5081E 05  0.6000E 05
DIS                    0.1481E 02  0.0         0.1100E 02  0.2348E 02  0.5408E 01  0.5000E 03

SOL   13  0.3804E-01   0.1039E 05  0.0         0.1018E 05  0.8525E 03  0.1577E 03  0.9188E 04  0.2078E 05  0.2997E 05
DIS                    0.9842E 02  0.0         0.1100E 02  0.8122E 03  0.2299E 03  0.5000E 03

SOL   14  0.1377E 00   0.6947E 03  0.0         0.6809E 03  0.5700E 02  0.1054E 02  0.6143E 03  0.1389E 04  0.2004E 04
DIS                    0.3434E 02  0.0         0.1100E 02  0.2207E 03  0.6155E 02  0.5000E 03

SOL   15  0.1444E-01   0.2080E 05  0.0         0.2039E 05  0.1707E 04  0.3157E 03  0.1839E 05  0.4161E 05  0.6000E 05
DIS                    0.3434E 02  0.0         0.1100E 02  0.2207E 03  0.6155E 02  0.5000E 03

SOL   16  0.1237E-01   0.6934E 02  0.0         0.6796E 02  0.5689E 01  0.1052E 01  0.6131E 02  0.1387E 03  0.2000E 03
DIS                    0.3434E 02  0.0         0.1100E 02  0.2207E 03  0.6155E 02  0.5000E 03

SOL    9  0.3196E 00   0.4611E 03  0.1895E 03  0.3598E 03  0.3772E 02  0.6976E 01  0.3246E 03  0.9223E 03  0.1247E 04
DIS                    0.2398E 02  0.2427E 02  0.1100E 02  0.1161E 03  0.3177E 02  0.5000E 03

         PERCENT H20     LB/DAY      LB/DAY        LB/DAY       LB/DAY       LB/DAY      LB/DAY        LB/DAY       LB/DAY
STATION     MGD          CARBON       BOD        NBIOCARBON    NITROGEN    PHOSPHORUS FIXED MATTER      VSS          TSS

SCL   18  0.7059E 02   0.2495E 04  0.0         0.2446E 04  0.2047E 03  0.3786E 02  0.0         0.4990E 04  0.7197E 04
DIS                     .5933E 00  0.0         0.1900E 00  0.3814E 01  0.1063E 01  0.8639E 01

SOL   20  0.1000E 02   0.1050E 03  0.1402E 03  0.3000E 02  0.1000E 02  0.2000E 01  0.3000E 02  0.2236E 03  0.2536E 03
DIS                    0.4300E 02  0.5984E 02  0.1100E 02  0.1910E 02  0.4000E 01  0.5000E 03
```

```
CASE NO.   1
                                                  ACTIVATED SLUDGE PROCESS CALCULATION
 NAME OF STRUCTURE                                                                                           JUNE 1967
                                CAPITAL COST     AMORTIZATION       OPERATING COST
                                   DOLLARS       DOLLARS/YEAR        DOLLARS/YEAR      AMORTIZATION PLUS OPERATING COST
                                                                                        DOLLARS/YEAR     CENTS/1000 GAL
 PRELIMINARY TREATMENT          0.9389E 05        0.6332E 04         0.1004E 05         0.1637E 05        0.4485E 00
 PRIMARY SETTLER                0.3149E 06        0.2124E 05         0.2469E 05         0.4593E 05        0.1258E 01
 AERATOR                        0.6776E 06        0.4569E 05         0.4519E 05         0.9088E 05        0.2490E 01
 AIR BLOWERS                    0.9861E 05        0.6650E 04         0.1783E 05         0.2448E 05        0.6707E 00
 FINAL SETTLER                  0.2940E 06        0.1983E 05         0.0                0.1983E 05        0.5432E 00
 SLUDGE RETURN PUMPS            0.3823E 05        0.2578E 04         0.0                0.2578E 04        0.7064E-01
 CONTROL HOUSE                  0.3036E 06        0.2048E 05         0.0                0.2048E 05        0.5610E 00
 SLUDGE THICKENER               0.1307E 06        0.8817E 04         0.0                0.8817E 04        0.2416E 00
 DIGESTER                       0.2880E 06        0.1942E 05         0.1226E 05         0.3168E 05        0.8678E 00
 SLUDGE ELUTRIATION             0.1408E 06        0.9497E 04         0.0                0.9497E 04        0.2602E 00
 VACUUM FILTRATION              0.2683E 06        0.1810E 05         0.4324E 05         0.6134E 05        0.1680E 01
 SLUDGE INCINERATION            0.4903E 06        0.3307E 05         0.2099E 05         0.5406E 05        0.1481E 01
 SLUDGE DRYING BEDS             0.0               0.0                0.0                0.0               0.0
 CHLORINATION                   0.4013E 05        0.2706E 04         0.1945E 05         0.2216E 05        0.6071E 00
 SITE DEVELOPMENT               0.4997E 05        0.3370E 04         0.0                0.3370E 04        0.9234E-01

 TOTAL CAPITAL COST             0.3229E 07
 TOTAL AMORTIZATION AND OPERATING COST                                                  0.4115E 06
 COST PER 1000 GALLONS                            0.5966E 01         0.5306E 01         0.1127E 02



    MLASS = 0.7550E 03      MLBSS = 0.2879E 03     MLNBSS = 0.6226E 03      MLDSS = 0.4691E 02       MLISS = 0.2844E 03
     MLSS = 0.2000E 04       VAER = 0.2313E 01       VNIT = 0.2941E 01     RETURN = 0.4656E 00         CWR = 0.4066E 00
   CARREM = 0.7864E 00     BODREM = 0.9018E 00     NITREM = 0.1103E 00    PHOSREM = 0.8573E-01         EFF = 0.3798E 00
     FRPS = 0.5000E 00       URPS = 0.4000E 03       URSS = 0.3000E 01       XRSS = 0.1242E-01
      GPS = 0.1375E 04        APS = 0.1501E 02        GSS = 0.2000E 04        AFS = 0.1510E 02
      GTH = 0.7500E 03       GSTH = 0.9000E 01       ATHM = 0.3335E 04        TRR = 0.9500E 00
       GE = 0.8000E 03        GES = 0.9000E 01         AE = 0.1583E 04        ERR = 0.7600E 00         WRE = 0.3000E 01
     TDIG = 0.3300E 02         TD = 0.1500E 02       VDIG = 0.1526E 03      FRDIG = 0.5910E 00
    CIDIG = 0.2695E 00      C2DIG = 0.8550E 03     CH4CFD = 0.9306E 05     CO2CFD = 0.4887E 05
      VFL = 0.4900E 01        TVF = 0.2380E 00        AVF = 0.4418E 03       FOOD = 0.1071E 03
     DEGC = 0.2000F 02         DO = 0.1000E 01     CAER20 = 0.1000E 01       CAER = 0.1000E 01
                           AEFF20 = 0.5000E-01       AEFF = 0.4537E-01
                             CNIT = 0.3215E 00       CEDR = 0.1247E 00
   AIRCFD = 0.8939E 07     CFPGAL = 0.8663E 00      BSIZE = 0.9311E 04
    FECL3 = 0.6227E 01     ALK(1) = 0.2423E 03    ALK(12) = 0.2423E 03   ALK(13) = 0.3058E 04   ALK(15) = 0.9463E 03
   ECF    1.00 2.00 1.20 1.50 2.00 2.00 1.00 1.50 2.00 1.50 1.00 1.00 1.00 1.00 1.00
   CENG = 0.7036E-01  CTRP = 0.1000E 00  CTGO = 0.1500E 00  CLAND = 0.2000E-01  CCR = 0.1340E 01
   CCI = 0.1130E 01        AF = 0.6744E-01
```